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**Proceedings: Sixth Symposium on the Transfer and
Utilization of Particulate Control Technology**

Volume 1

Holly Rossi, Compiler

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P. O. Box 13000
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EPA Contract 68-02-3994, Task 46**

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16. ABSTRACT These proceedings document the Sixth Symposium on the Transfer and Utilization of Particulate Control Technology, in New Orleans, LA, February 25-28, 1986. The symposium, the second of its kind to be sponsored jointly by EPA and EPRI, was devoted to discussions of particulate emissions control. Its objectives were to promote the transfer of results of particulate control research to potential users of technologies, and to provide a medium for exchange of ideas between researchers in the fields of particulate control development and application. The proceedings contain 84 papers presented by EPA and other federal and state agency representatives, EPRI staff members, representatives of utility companies, equipment and process suppliers, university representatives, and research and development companies. Additionally, individuals from Australia, Canada, Denmark, Great Britain, Italy, Japan, and South Africa presented information on worldwide technological developments. Electrostatic precipitators and fabric filters were the major topics discussed. Other topics of interest included system operation and maintenance, fugitive emission control, integrated control processes (e.g., SO ₂ and particulate control), and advanced energy applications (e.g., high temperature applications).		
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a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Pollution Processing	Pollution Control	13B 13H
Particles Leakage	Stationary Sources	14G
Electrostatic Precipitators	Particulate	13I
Filters Sulfur Dioxide	Fabric Filters	11E 07B
Fabrics Nitrogen Oxides	Fugitive Emissions	15E
Maintenance		
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PREFACE

The Sixth Symposium on the Transfer and Utilization of Particulate Control Technology was held in New Orleans, Louisiana, February 25 through 28, 1986. This Symposium, the second of its kind to be jointly sponsored by EPA and EPRI, was devoted to the discussion of particulate emissions control. The Symposium's objectives were to promote the transfer of results of particulate control research to potential users of technologies, and to provide a medium for exchange of ideas among researchers in the fields of particulate control development and application.

The Symposium proceedings contain 84 papers presented by EPA and other Federal and State agency representatives, EPRI staff members, representatives of utility companies, equipment and process suppliers, university representatives, and research and development companies. Additionally, individuals from Australia, Canada, Denmark, Great Britain, Italy, Japan, and South Africa presented information on worldwide technological developments. Electrostatic precipitators and fabric filters were the major topics discussed during the Symposium. Other topics of interest included system operation and maintenance, fugitive emission control, integrated control processes (e.g., SO_2 and particulate control or NO_x and particulate control), and advanced energy applications (e.g., high temperature or pressure applications).

Symposium cochairmen were Ralph F. Altman, Project Manager in the Air Quality Control Program of the EPRI's Coal Combustion Systems Division; and Dale L. Harmon, Project Officer in the Particulate Technology Branch of the EPA's Air and Energy Engineering Research Laboratory. Participants in the introductory plenary session were John S. Maulbetsch, Program Manager in the Air Quality Control Program of the EPRI's Coal Combustion Systems Division; Norman Plaks, Chief of the Particulate Technology Branch of the EPA's Air and Energy Engineering Laboratory; and keynote speaker Grady B. Nichols, Southern Research Institute.

ABSTRACT

These Proceedings document the Sixth Symposium on the Transfer and Utilization of Particulate Control Technology, held February 25 through 28, 1986, in New Orleans, Louisiana. The Symposium was sponsored by the U. S. EPA's Air and Energy Engineering Research Laboratory in Research Triangle Park, North Carolina, and EPRI's Coal Combustion Systems Division, in Palo Alto, California.

The objectives of the Symposium were to provide an opportunity for the exchange of knowledge and to stimulate new ideas in the field of particulate control research with the goal of extending the technology and aiding its diffusion among designers, users, and educators. Fabric filters and electrostatic precipitators were the major topics, but novel concepts and advanced technologies were also explored. Topics discussed in each day's sessions are listed below:

Tuesday, February 25

- Plenary Session
- ESP: Hot-Side Precipitator
- FF: Advanced Concepts I
- Scrubbers
- ESP: Performance Enhancement I
- FF: Advanced Concepts II
- Opacity/Measurements

Wednesday, February 26

- ESP: Performance Enhancement II
- FF: Practical Considerations
- Advanced Energy Applications I
- ESP: Performance Estimating (Modeling)
- FF: Pilot-Scale Studies
- Advanced Energy Applications II

Thursday, February 27

- ESP: Advanced Technology
- Operation and Maintenance

- Integrated Control Processes I
- ESP: Fundamentals I
- ESP: Controls and Energy Consumption
- Integrated Control Processes II

Friday, February 28

- ESP: Fundamentals II
- ESP: Design Considerations
- Fugitive Emissions

Volume 1 of the Proceedings contains 26 papers presented in the Scrubbers, Opacity/Measurements, Advanced Energy Applications I and II, and Integrated Control Processes I and II Sessions.

Volume 2 of the Proceedings contains 35 papers on electrostatic precipitator technology presented in the Hot-Side Precipitator, Performance Enhancement I and II, Performance Estimating (Modeling), Advanced Technology, Fundamentals I and II, Control and Energy Consumption, and Design Considerations Sessions.

Volume 3 of the Proceedings contains 23 papers presented in the FF: Advanced Concepts I and II, FF: Practical Considerations, FF: Pilot-Scale Studies, Operation and Maintenance, and Fugitive Emissions Sessions.

Seven papers presented at the Symposium (nos. 3, 37, 41, 50, 51, 55, and 76 in the Symposium program) were not submitted for publication in the Proceedings.

ACKNOWLEDGEMENT

The assistance of Holly Rossi, Radian Corporation, in compiling these Proceedings is gratefully acknowledged.

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VOLUME 1

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PLENARY SESSION

Ralph F. Altman, Chairman
Electric Power Research Institute
Chattanooga, TN

KEYNOTE ADDRESS

PARTICULATE MATTER CONTROL - THE RAPIDLY ADVANCING TECHNOLOGY

by: Grady B. Nichols
Director, Environmental Sciences Research
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It is a pleasure and an honor to be allowed to deliver the keynote address to this Sixth Symposium on the Transfer and Utilization of Particulate Control Technology.

I will begin my presentation today by acknowledging the two organizations that are responsible for conducting this meeting - EPA and EPRI. These two organizations are the fundamental driving forces that provide an opportunity for us in the research community to conduct the research that we are able to share in meetings such as this. Before EPA and EPRI existed, such research was primarily conducted for the in-house use of various particulate matter control companies. With the formation of what is now EPA and a few years later the Electric Power Research Institute, funds became available to expand the research areas to include groups outside those manufacturing the control devices. This at least increased the number of people available to address the pressing needs for research related to the control of particulate emissions into the atmosphere. I express my personal appreciation for this support and feel sure that I can say that a similar appreciation is felt by the entire community that is represented here.

Some of the most important recent advances in particulate control technology can be traced directly or indirectly to one of the organizations sponsoring this symposium. In instances where the research topic was not initially related to either EPA or EPRI, their efforts have been responsible for bring the technology to our attention. Those of us who have been involved in this research have either been funded to do the work or allowed to visit with those organizations around the world that were actually conducting the research. Our contacts in Australia and Japan were initially made possible through the contributions of EPA and EPRI; the technology transfer between the U.S. and Soviet Union, as well as recently with the Peoples Republic of China, were also made possible through their efforts.

The continuing series of symposia provide an opportunity for those research organizations in Europe to share their research results with the rest of the world. I feel that these efforts have been a significant factor in the reduction of particulate emissions on a global basis, making their existence very worthwhile.

I will turn now to the specific topics of my presentation. The topics that will be discussed will include new developments in electrostatic precipitator research and in fabric filter research and will end with some discussions related to the inhalable particle standards and acid rain legislation.

ELECTROSTATIC PRECIPITATION

Power Supplies and Controls

The more significant developments in the area of electrostatic precipitation include the new microprocessor controls and pulse-intermittent energization. The microprocessor controls add a significant degree of sophistication to the control of precipitator power supplies. Whereas, the older controls are equipped with current and voltage limits and spark rate controls, the microprocessor controls, for most purposes, are limited only by the imagination of the programmer. In addition to the older control functions, the new system can sense changes in the shape of the voltage-current curves and thus avoid back corona, can be controlled by opacity or other variables for energy management, or can be programmed to skip a given number of input cycles to operate with what is termed "intermittent energization." Even in those instances when high resistivity is not a problem, the energy management concerns of precipitator operation are sufficient to justify the use of these new control systems.

This technology is especially useful to installations that operate with a variable coal source or have cyclic operation. If the installation is expected to operate over its lifetime with wide variations in resistivity, particle size and loading, the unit can be designed for the worst-case conditions and then be programmed to provide an adequate degree of control, while conserving electrical power. This flexibility in control is extremely useful to a modern installation.

Both pulse and intermittent energization appear to have a place in the operation of modern-day precipitators. Even though pulse energization is not a new concept (Dr. White investigated and reported on this in the 1950's), progress in modern-day electronics has allowed the technology to mature. Initially, it was conjectured that pulse energization would provide a means for circumventing the problems of high resistivity by effectively charging the particles to a high value while maintaining a low value of current density to avoid back corona; this high level of charging was not borne out in practice.

The principal advantage of pulse energization is for the collection of high resistivity particles. If a very high voltage pulse is applied to a typical electrostatic precipitator field, corona is generated over essentially the entire length of the wire. This high voltage pulse leads to a much more uniform current density through the dust layer that exists on the collection electrodes. If an attempt is made to operate a conventional system at a comparable voltage (negative corona), the current density distribution in the collection electrode will exhibit a large spacial variation with the tendency to develop back corona at those areas with higher current density. The pulse energization will cause current to flow from essentially the entire surface of

the wire rather than from several isolated spots for a given average current density. This factor is the primary advantage of pulsing over conventional excitation.

Intermittent energization is a method of powering the precipitator somewhere between conventional and pulse methods. In this approach, the power supply is programmed to omit a set number of cycles of the input waveforms and to pass one cycle. The number of cycles omitted is preselected based on the average current density desired for a given resistivity. Since the back corona formation is related to the average current density through the dust layer, the current density and peak voltage can almost be selected independently. The operating voltage can be driven well up the voltage current curve to yield a high value of voltages, while the average current density can be held below that which causes back corona. The high voltage operation leads to a more uniform instantaneous current density (similar to pulse energization), while the average current density is maintained at a low value.

Even though the intermittent energization is qualitatively not as good as pulse energization, the system cost is much less than a pulse system. The microprocessor control can be programmed to operate in an intermittent mode, while a second series of power supplies is required for pulse excitation.

Electrode Modification

The basic electrostatic precipitator which has evolved has a spacing between collection electrodes on the order of 12 inches and a corona electrode with a diameter of about 0.1 inch. These dimensions were established somewhat because of practical availability of power supplies that could deliver operating secondary voltages on the order of 50 kilovolts at a current density of about 50 microamperes per square foot of collecting area. These design dimensions are now being reevaluated on a worldwide basis. We see plate spacings increasing to the range of 20 inches, as well as some limited evaluations of corona electrodes with diameters approaching one-half inch. These changes are of interest now because of the availability of power supplied that can reliably operate in the 100-kilovolt range.

Reports have been available for some time indicating that an increase in plate spacing causes an increase in performance such that the same collection efficiency can be maintained with a proportional decrease in the required collection electrode area. This reduces the amount of material required for the construction of the precipitator. The additional space charge from the charged particulate and ions increases the collecting electric field to compensate for the reduction in plate area.

A second factor that is important in the wider plate spacing is a reduced sensitivity to high resistivity particulate matter. The wider spacing leads to a more nearly uniform current density distribution on the collecting electrode, thus the precipitator becomes less sensitive to the high resistivity tendency to form back corona. There is some evidence that the increased electric field at the dust layer surface is also a factor in avoiding back corona. Therefore, I see some promise in this study and in the use of wider plate spacings in the new precipitators.

The interest in new corona electrode systems is in developing a system that will increase the value of the collecting electric field while maintaining relatively low values of current density. Since back corona development is related to the average operating current density (as well as the resistivity), if an increased collecting electric field can be formed at a lower value of current density, the formation of back corona can sometimes be avoided. The approach to obtaining these desirable characteristics include substituting larger diameter round wires as well as a number of new individual designs that have been tried for the rigid discharge electrodes.

These new looks at the basic design dimensions of the electrostatic precipitator are expected to yield significant results in the not-too-distant future. Operating at significantly higher voltages with significant reduction in the current density will not only improve the performance of the precipitator, but will in many instances lead to a reduction in the electrical power required to operate the system.

Another new development related to collection electrodes is the rotating collection electrode system equipped with a rotating brush to clean the electrodes. Several precipitators in Japan are equipped with this design. When this system was first considered, there was some skepticism concerning the practicality of the approach for full-scale operating units. Areas of concern were: (1) mechanical problems associated with the moving electrode system in the dirty gas stream, and (2) maintaining an electrode surface sufficiently clean to avoid back corona. However, operating experience has indicated that it is technically feasible and it seems to be merely a question of economics as to whether you select this system, sulfur trioxide conditioning or a larger precipitator to meet the emission needs of an installation.

Prechargers

An approach to the collection of high resistivity particulate matter involves the development of multistage precipitators. The philosophy is that particle charging can be effectively carried out in a very short section of the precipitator while collection can be handled in the larger portion where high electric fields and low current densities can be maintained. We have heard discussions of various approaches to the development of these prechargers at several of these symposia. In general, it is thought that a high current density precharger can be built that avoids back corona. Since the precharger is very small, it is reasonable to use exotic approaches in it that might be prohibitively expensive in an entire full scale precipitator. Chargers used in conjunction with electrode modification continue to look promising.

Model Development

There has been a significant amount of work put into the development of mathematical models of electrostatic precipitation. In the early 1900's, Deutsch prepared a theoretical development model with a basic set of assumptions of the behavior of a precipitator. In the same time frame, Anderson reviewed test data that were taken from a number of precipitator performance tests. These two studies led to the development of a single parameter model, now known as the Deutsch-Anderson (D.A.) equation. This is

an excellent mathematical model when applied under the original basic assumptions of operating conditions with all particles with the same migration velocity, no gas leakage, no reentrainment, and uniform particle distribution across lane of the precipitator.

White later extended the theory to include a wide range of particles that had a log normal size distribution. The selection of a log normal distribution allows the D.A. equation to be integrated over the particle size range to develop a two parameter model that would apply to the entire precipitator. Matts and Ohnfeldt independently reviewed performance data for a wide range of precipitator operation and also developed a two-parameter model, similar in form to that developed theoretically by White. This model, currently referred to as the W_k model, is widely used in the industry.

In the later 1960's, Southern Research Institute developed and is continually updating the first computer systems model that was based on first principles for charging and collecting the particles. The Deutsch-Anderson equation is applied to narrow size bands of particles and short increments of length in the precipitator and later coupled with empirical relationships for sneakeage and reentrainment. This model is used extensively in the industry.

Roy Bickelhaupt developed a resistivity prediction model for fly ash based on the chemical composition of the fly ash and flue gas. The model takes into account the sulfur trioxide concentration in the gas stream. The greatest uncertainty in the model is the estimate of the amount of sulfur trioxide that is present in the gas stream. Stoichiometric calculations are made for the sulfur dioxide and a percentage of this is used as the sulfur trioxide. When measured values for sulfur-trioxide are used, the prediction is extremely good. This model needs to be expanded to include fly ash that also contains sorbents for sulfur dioxide removal. Currently, this research is under way.

More recently, Self and Mitchener of Stanford and McLean of Australia and Sparks (of EPA) have conducted detailed studies of the precipitation process. These models are intended to look more deeply into the behavior of the particles and ions in the precipitation process.

The Stanford work is intended to look into the possibility of improving the performance of a precipitator by modifying the gas-flow turbulence. If one compares the performance of a laminar flow precipitator with one that has fully developed turbulence, there is a significant difference. In a characteristic length, a laminar flow device would theoretically collect 100 percent of the particles while a Deutsch model would predict about 63 percent ($1 - 1/e$). Thus, if one could approach a laminar flow condition, a significantly smaller precipitator would be required.

There are two serious drawbacks to this development. First, the physical dimensions of a laminar-flow precipitator are such that it is difficult to design and build. Soda straw types of dimensions are used in some instrumentation systems, and obstructions such as corona electrodes would be expected to generate some turbulence. Second, the emissions from a high efficiency precipitator may well be dominated by reentrainment. The rapping reentrainment from a conventional precipitator is related to the amount of

material collected in the outlet field. In a laminar-flow precipitator, a significantly greater percentage of the material would be subject to reentrainment.

Even with the above consideration, the potential for improvement in performance, as well as the possibility of developing a better understanding of the precipitation process, justify the continuation of this research.

Work related to the development of a detailed understanding of the corona current distribution from a corona wire is providing new information needed for the development of a model. The development of back corona with a collected layer of high resistivity material is related to the local current density through the dust layer. These studies, based on a negative corona system, have shown that the current from each individual corona tuft can be described as a saturated current distribution that is well-defined (Warburg distribution). As the voltage is increased on a wire, additional tufts are formed, each with its pattern. As the tuft density on the wire increases, the patterns from adjacent tufts attempt to intersect. In actuality, the patterns merely truncate at the intersection point. This provides a convenient means for modeling the current distribution in a precipitator.

This modeling effort needs to be extended to include back corona. If back corona can be successfully modeled, then an additional understanding of the formation and, perhaps, prevention of back corona can be developed.

Conditioning Studies

The two most significant fundamental factors that limit the performance of an electrostatic precipitator are the particle size distribution and electrical resistivity of the particulate matter. The collection efficiency as a function of particle size exhibits a minimum in the particle diameter range on the order of 0.5 micrometers. Therefore, the greater the amount of material contained in this size range, the more difficult the collection problem. A second related factor is the electrical space charge that the ultrafine particles contribute, which leads to a current suppression in the inlet field.

High electrical resistivity limits the performance of the precipitator by causing either sparking at reduced current densities (and voltage) or back ionization. The corona current that flows in the interelectrode space must flow through the collected dust layer in completing the circuit back to the power supply. This current flow through the layer establishes an electric field in the layer proportional to both the current density and the electrical resistivity of the layer. At some value of electric field, sparking is initiated by localized breakdown of the layer. This local spark-through behaves as a point on the collection electrode that leads to a sparkover to the corona wire at an operating voltage that is less than would be required to spark for a smooth collection electrode.

As the resistivity is increased, for a given current density, the electrical breakdown of the gas spaces within the dust layer occurs at voltages less than are required to cause a sparkover. This leads to a general electrical breakdown in the collected layer, which is termed back ionization or back corona. When this occurs, positive ions from the plate are injected

back into the interelectrode space where the desirable negative ions from the corona process are and cause a net decrease in the space charge, causing a further increase in current and a reduction in the electric field that acts to charge and collect particles.

Conditioning studies in the past have sought means to modify the particle size distribution as well as the resistivity of the particulate matter. Attempts to develop particle agglomeration through sonic energy or modification of the cohesivity of the particles have been essentially failures. It is not surprising that these efforts have failed, when one considers that in a power station's flue gas, the particles are separated by about 50 mean diameters, and that the energy required to bring about a significant number of collisions between particles will be very large.

Particle agglomeration does, in fact, occur within the collected dust layer when the particle cohesivity is increased. This increased cohesivity causes the reentrained particles to be dispersed with a larger size distribution, thus minimizing rapping reentrainment. An example of this increased cohesivity is ammonia injection in the presence of sulfur trioxide, which chemically combines to form ammonium sulfate or bisulfate. A similar reaction occurs when both ammonia and sulfur trioxide are jointly injected.

The most successful conditioning method has been the injection of sulfur trioxide to reduce the resistivity of fly ash materials. The mechanism is well understood and many successful installations are currently operating. A new method of sulfur trioxide conditioning is now under study. The thermal decomposition of ammonium sulfate is conducted in a side stream, and this gas stream is then injected into the main flue gas stream below the air heater. Injection just ahead of the precipitator avoids fouling of the air heater, which results from thermal decomposition on the high-temperature side of the heater. The thermal decomposition study is aimed at developing an alternative sulfur trioxide conditioning system to the current system of the catalytic oxidation of sulfur dioxide to trioxide.

Rapping Reentrainment

One aspect of electrostatic precipitation that must be addressed in greater detail is rapping reentrainment. While the fundamental factors of particulate collection are well understood and can be adequately modeled from first principles, rapping reentrainment is largely understood in terms of empirical relationships. It may well be that rapping reentrainment is so complicated and site-specific that an effective model cannot be developed. However, we will never know this until a very detailed research project is conducted to verify this. I challenge this group of people to start on this study now.

Wayne Sproull conducted an empirical study in the early 1960's where he investigated the rapping accelerations required to dislodge a variety of dusts. Southern Research conducted limited pilot-scale and full-scale studies in the 1970's to determine the amount of emissions that is related to rapping reentrainment for coal-fired furnaces. These empirical studies could be used in formulating a detailed powder technology study of the requirements and mathematical modeling of this phenomena.

Electron Beam and Free Electron Charging

The current research into the use of high-energy electron beams for the charging of particles and the promotion of sulfur oxide removal is providing some new results. I believe that the evidence supports the ability to provide a significant charge density in the vicinity of this electron beam when the experiments have been conducted. However, I am very skeptical of this technology actually being brought to commercial practice within the next couple of decades. It is one thing to be able to provide a small volume beam of high-energy electrons in an adequately shielded, laboratory-scale electrostatic precipitator. It is quite another to provide such a system for use in an industrial application, where a number of precipitators, each with a cross-sectional area of the inlet ducts, approaching 1,000 square feet or more. Even if the system could be built, the very large space charge of charged particulate matter would be expected to develop electrical breakdowns in the inlet plenums to the precipitator unless very effective electrical shielding could be devised. In my opinion, this is very interesting university research, but has little promise for actually solving an industrial precipitator problem in the near future.

FABRIC FILTRATION

Fundamental research on fabric filtration has not yet matured to the same stage as electrostatic precipitation. In fact, I consider it to be at about the stage that ESP research was in the early 1970's. This is, in part, related to the general consensus that the fabric filter would attain a collection efficiency near 100 percent with any dust. The only real concern was what pressure drop would develop across the bag system and what were the bag replacement requirements. In many cases these were the only problems, but we are now seeing some exceptions to this in the utility industry. One significant exception that comes to mind is the Monticello Station in Texas. In this installation, the fly ash characteristics are such that the individual particles are very smooth and tend to bleed through the fabric-dustcake composite and yield excessive emissions. A rather detailed study of this installation led to the conclusion that ammonia conditioning was required to reduce the emissions. There is now a growing recognition that different coals yield fly ashes with diverse filtration properties. This variability leads to studies attempting to better define such things as operating gas to cloth ratios, cleaning modes, and fabrics for each fuel.

There are several areas of significant research that are currently active and a few more that should be started. The conventional practice of reverse-gas-cleaned bags in the utility industry is now being questioned. Studies of sonic energy assist and shake-deflate cleaning are providing alternative approaches to reduce the operating pressure drop of these installations. These studies are related to the optimization of frequency, amplitude, duty cycle, and fabric for either reverse gas or shake-deflate cleaning for types of fuel and ash. If these studies show that pressure drop can be reduced with no significant decrease in bag life, the use of fabric filtration can certainly be expanded and become more competitive with electrostatic precipitation.

Alternative Fabric Studies

A necessary area of research is a study of alternative fabric materials for the baghouse. The current practice centers on the use of coated fiberglass bags in the United States. Most coal-fired utility experience involves as a typical construction a 3 x 1 twill fiberglass with either a teflon, tricoat, or acid resistant finish. I feel that this choice is dictated by the specification of surviving a temperature in the range of 500°F. I suspect that this is an unnecessarily strict specification, and it places a severe restriction on the types of candidate materials for use in the baghouse. If the intent is to prevent damage to the baghouse in the event of a failure in the air heater, I do not think that it will. A failure in the heater would usually result in a temperature excursion to 650 - 700°F, depending on the installation. If this specification were relaxed to about 350°F, a greater variety of materials would be considered, including some of the acrylics that are used in Australia. Several additional fabrics have recently become available in the U.S. for trial and evaluation.

A new study group has just been formed to further develop the technology of fabric filtration. The need for the formation of this study group was recognized in a Fabrics and Filterbag Workshop held in Atlanta by EPRI in the Fall of 1985. Dan Giovanni of Electric Power Technologies is coordinating the activities of this fabric and filterbag manufacturers advisory group. In addition to the manufacturers, representative of the electric utility industry are included. The first meeting was held in Denver on February 13, 1986, at the EPRI Arapahoe test facility. Participation activities in this group will include:

- Formulation of Experimental Test Plans for Fabric Studies in the EPRI Fabric Filter Pilot Plants;
- Standardization of Fabric Testing Methodology;
- Specification, Procurement, and Fabrication of Test Filterbags;
- Preparation of a "Primer" on Filterbags for Utility Baghouses;
- Standardization of Filterbag Procurement Procedures; and
- Interpretation and Dissemination of Research Results.

If you have an interest in following the results of this study group, contact Dan Giovanni, Bob Carr, or Wallace Smith of Southern Research. I expect a significant amount of progress from this study group that is useful to the utility industry as well as others.

Powder Technology and Model Development

Another aspect of fabric filter research that is needed is to develop a better understanding of the characteristics of the dust-fabric interface and the dust layer (powder technology) itself. A definition of the dust behavior, including shear strength, adhesivity, and cohesivity, and the factors that influence these relationships would provide insight into what factors are important in developing an effective filtration system.

Another aspect of needed research, which is related to the above study, is the development of a predictive model for fabric filtration. An effective predictive model would allow one to take better advantage of results of research and to develop more confidence in the design of a collection system for a new installation, similar to that which is now available for electrostatic precipitators.

Electrostatic Augmentation

Several studies of electrostatic augmentation by either precharging the dust particles, adding an electrostatic field to the fabric filter system, or a combination of both charging and electrostatic collection have shown a consistent reduction in pressure drop. For the special case where the particles are charged and an electrostatic collection field is added, the collecting process is significantly improved. Since the electrical velocity of the charged particles is comparable to the gas velocity through the bag, the particles will essentially follow a deterministic path such that a significant portion of the fabric will be essentially free of particles, while collection occurs primarily in the vicinity of the grounded electrodes. While this does provide a high collection efficiency at a much reduced pressure drop, it remains to be seen whether the costs of adding the electrical complication to a fabric filter system, which is in itself expected to cost essentially the same as an electrostatic precipitator, is warranted. There is also the question about what the influence of high resistivity will be on this system.

EXPECTED CHANGES THAT RELATE TO PARTICULATE CONTROL

There are two significant changes or developments that we in the particulate control community should be aware of: (1) the expected establishment of an inhalable particulate standard referred to as PM_{10} and (2) acid rain legislation. Both of these actions are expected to require an improvement in the collection efficiency of control devices.

The PM_{10} standard will probably be established in the Fall of 1986. The current₃ considerations are for a primary standard in the range of 50 to 65 $\mu g/m^3$. This represents the concentration of particulate matter that is smaller than 10 micrometers in diameter. If the PM_{10} standard is established at the lower end of the range, it will probably require a reduction in the allowable emissions from those now established in many industries. I do not have the data to state what sources will require additional control levels, but it is probable that some combustion sources, such as coal-fired utilities, may require additional control, while others, such as a silicon metal furnace in the ferroalloy industry, will almost surely require additional control. When these new standards are established, we will likely have an opportunity to implement some of the new technology that we will have discussed at this symposium.

A second, and somewhat more stringent, problem is the expectation that acid rain reduction legislation will be passed. There is a considerable amount of research currently aimed at the reduction of sulfur oxides. Such acronyms as LIMB and DSEC, as well as spray dryers and the related ESO_x modification to electrostatic precipitators, are familiar research topics

related to the control of acid rain precursors. The discussion of these items generally center around their ability to reduce sulfur and nitrogen oxide emissions.

In the LIMB and DSEC technologies, calcium-based sorbent materials are injected into a coal-fired furnace in a temperature range of 2,000°F. These finely divided particles are quickly calcined to calcium oxide which then reacts with sulfur dioxide and oxygen to form calcium sulfate. Since the calcium utilizations obtained at typical calcium to sulfur ratios of 2 to 1 are on the order of 25 percent, furnace injection of these sorbents will result in large increases in particulate matter loadings to downstream collection equipment.

The ESO process differs significantly from the furnace injection and is more closely^x related to spray dryer technology. The chemical reactant is dissolved into water and delivered as a very fine mist into the inlet region of an electrostatic precipitator at a temperature of about 300°F. Reaction occurs between the droplets and sulfur dioxide while the water in the droplets is evaporated. The native sulfur trioxide has already interacted with the fly ash to effectively condition the ash while the evaporation of the water reduces the gas temperature to further reduce the resistivity. This further reduction in resistivity is hoped to compensate for the increased mass loading into the precipitator. A secondary effect is to slightly reduce the gas volume.

A major concern for this group, which is frequently neglected in the considerations, is the effect that sulfur oxide control will have on particulate control. In all instances other than wet scrubbers, the implementation of acid rain control will place a significantly higher burden on the particulate control device.

The first factor to consider is that the addition of a chemical to react with sulfur oxides, with the resulting reaction products, essentially doubles the mass loading into the particulate control device. Therefore, if we start with a particulate control device that is designed for an intermediate-to-high-sulfur installation that is currently meeting the particulate matter emission standards, the installation will no longer meet these standards after SO₂ controls are added. Therefore, the particulate control device must be upgraded to maintain compliance.

A second and even more important factor is that of the options available; essentially all of them will be further detrimental to the performance of electrostatic precipitators. The injection of calcium or other alkali chemicals to remove the sulfur dioxide will result in the removal of essentially all of the sulfur trioxide from the gas stream. Switching to low-sulfur coal and coal cleaning yields a similar result. Therefore, a precipitator that is designed with a specific collection electrode area of 300 - 400 square feet per thousand cubic feet per minute of gas for a resistivity in the 1,0¹⁰ ohm-centimeter range will be totally inadequate for resistivities in the 1,0¹² ohm-centimeter range. When the acid rain controls are implemented, it will become necessary to modify the resistivity through conditioning and moisture and temperature modification. In some instances, upgrading or increasing the size of the control device to handle the increased mass loadings may be required. Therefore, the particulate control community

must be prepared to handle these more difficult particles that will result from the application of sulfur oxide control technology to existing plants that are currently in compliance. This new legislation also will require the application of some of the technology that we will hear discussed at this meeting.

Several items that come to mind for the upgrading of these existing units include the modification to the current electrode systems, conversion to multistage units with charging sections added, and new approaches to energization. In other cases, it may be necessary to retrofit fabric filters in conjunction with or to replace the existing particulate matter control devices.

CONCLUSION

I look forward to hearing the results of some very exciting research topics that are included in this symposium. I challenge you to listen carefully and participate in the discussions of these papers. Please make full use of the question period and continue the discussions during the break periods. The presenters are usually very enthusiastic about discussing their research efforts and would welcome discussion that may lead them into a better understanding of their research results as well as into new areas of related research. We are fortunate to be able to participate in this very exciting period of research related to the control of particulate matter into the atmosphere.

Session C-1: SCRUBBERS

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DEVELOPMENT OF IMPROVED SINGLE DROP COLLECTION EFFICIENCY CORRELATIONS
FOR MICROCOMPUTER MODELLING OF VENTURI SCRUBBER PERFORMANCE

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ABSTRACT

The overall collection efficiencies of high energy venturi scrubbers depend on a number of particulate matter collection mechanisms operating simultaneously. The predominant modes for particle capture are inertial impaction and interception by liquid drops acting as spherical collectors. With available models, prediction of single drop collection efficiencies has been based on potential or Stokes flow fields around the individual drops. However, experience shows that for industrial scale venturi scrubbers, flow conditions are intermediate between potential and viscous regions. Current models of venturi scrubber performance assume the flow fields to be strictly potential. On this basis, predicted overall collection efficiencies overestimate the experimental values. This investigation establishes the Reynolds number range over which the transition from potential to viscous flow will occur. Correlations for predicting realistic single drop collection efficiencies are provided for use in improved performance models.

INTRODUCTION

The objective of this work is to evaluate particle collection efficiencies of single water droplets for the intermediate Reynolds number range by considering the principal collection mechanisms to be inertial impaction and interception. There is a considerable body of literature describing particle deposition on single spherical targets assuming viscous and inviscid flow conditions (1-7). The characteristics of single drop/particle interactions are important when modelling the performance of wet gas cleaning equipment.

The flow fields around collector droplets generated during typical venturi scrubber operations are not strictly potential or viscous in nature (8). They depend on the throat length, throat gas velocity and location of liquid injection. Since current venturi scrubber performance models assume the flow fields to be entirely potential, they overestimate overall particulate matter collection efficiencies. Placek (8) has emphasized the need to develop capabilities of predicting single drop collection efficiencies for the intermediate Reynolds number range, however relatively little has been done in this area.

THEORETICAL MODEL

A single droplet collection efficiency can be defined in several different ways. In this work, it has been taken as the ratio of the number of particles predicted to strike the collector to the number of particles flowing through the projected area of the collector. Particle collection will occur from within an area defined by trajectories that would be traced by point masses that just graze the collector as shown in Figure 1. If y'_0 is the initial displacement of the limiting trajectory from the collector centreline in the undistributed flow region, the droplet collection efficiency is given by:

$$E = \left(\frac{y'_0}{r_c} \right) \quad (1)$$

where r_c is the radius of the collector.

A particle trajectory is determined from the equations of motion which are derived by performing a force balance on a particle that is interacting with the flow field of a spherical collector. In their dimensionless forms these equations reduce to:

$$\frac{d^2x}{dT^2} = \frac{1}{K} (U_x - V_x) \gamma_x \quad (2)$$

$$\frac{d^2Y}{dT^2} = \frac{1}{K} (U_y - V_y) \gamma_y \quad (3)$$

where γ_x and γ_y are correction factors that account for non-Stokesian behavior of particles when the Reynolds number $Re_p > 0.01$. The values of γ_x and γ_y have been expressed in the form:

$$\gamma = C_D \frac{Re_p}{24} = 1 + \frac{Re_p}{109} + 0.15 Re_p^{0.6} \quad (4)$$

by using the expression for the drag coefficient given by Dickinson and Marshall (9).

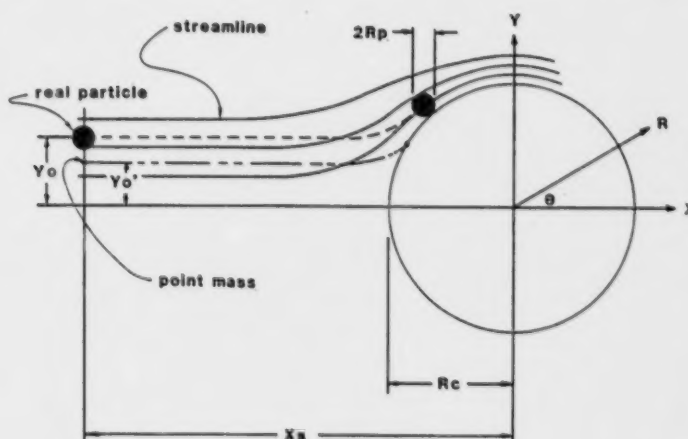


Figure 1. Schematic illustration of limiting trajectories.

Since the fluid flow field around the collector is not strictly viscous or inviscid, the Navier-Stokes momentum equations must be solved by considering both viscous and inertial force terms. Many researchers have attempted to develop approximate solutions for the intermediate flow region with moderate success. In this work, the equation developed by Hamielec and Johnson (10) was used for Reynolds numbers ranging between 10 and 80. For Reynolds numbers less than 0.4, the equation of Tomotika and Aoi (11) was employed.

The Hamielec (10) equation is given in terms of a stream function according to:

$$\psi = \left(\frac{R^2}{2} + \frac{A_1}{R} + \frac{A_2}{R^2} + \frac{A_3}{R^3} + \frac{A_4}{R^4} \right) \sin^2 \theta + \left(\frac{B_1}{R} + \frac{B_2}{R^2} + \frac{B_3}{R^3} + \frac{B_4}{R^4} \right) \sin^2 \theta \cos \theta \quad (5)$$

where

$$A_2 = \left(\frac{-125-120\beta}{60+29\beta} \right) + \left(\frac{-140-75\beta}{60+29\beta} \right) A_1 \quad (6)$$

$$A_3 = \left(\frac{135+153\beta}{60+29\beta} \right) + \left(\frac{108+63\beta}{60+29\beta} \right) A_1 \quad (7)$$

$$A_4 = \left(\frac{-40-47.5\beta}{60+29\beta} \right) + \left(\frac{-28-17\beta}{60+29\beta} \right) A_1 \quad (8)$$

$$B_2 = \left(\frac{-140-69\beta}{60+27\beta} \right) B_1 \quad (9)$$

$$B_3 = \left(\frac{108+57\beta}{60+27\beta} \right) B_1 \quad (10)$$

$$B_4 = \left(\frac{-28-15\beta}{60+27\beta} \right) B_1 \quad (11)$$

Values of A_1 and B_1 were provided (10) in graphical form as functions of β (viscosity ratio of the dispersed phase to the continuous phase) and Reynolds number. For the air/water system the value of β is large. Consequently, A_1 and B_1 have been evaluated from the graphical representations in the forms:

$$A_1 = 1.841 \left(\frac{Re_c}{100} \right)^2 - 4.493 \quad (12)$$

$$B_1 = -0.1 Re_c \quad (13)$$

The Tomotika and Aoi (11) equation is given by:

$$\psi = - \left[- \left(R - \frac{1}{R} \right) - \frac{16+3Re_c}{32} \left(R^2 - \frac{1}{R} \right) + \frac{3Re_c}{32} \left(R^2 - \frac{1}{R^2} \right) \cos \theta \right] \sin^2 \theta \quad (14)$$

The dimensionless velocity components can be calculated from Equations 5 and 14 according to:

$$U_R = \frac{1}{R^2 \sin \theta} \frac{\partial \psi}{\partial \theta} \quad (15)$$

$$U_\theta = \frac{-1}{R \sin \theta} \frac{\partial \psi}{\partial R} \quad (16)$$

where

$$U_x = U_R \cos \theta - U_\theta \sin \theta \quad (17)$$

$$U_y = U_\theta \cos \theta + U_R \sin \theta \quad (18)$$

PROBLEM SOLUTION

To determine the collection efficiency of a spherical collector, Equations 2 and 3 must be integrated to yield the limiting trajectory of a point mass particle. These equations were integrated numerically using a variable step-size algorithm to accommodate the large change in the radius of curvature of the trajectory as the particle approaches the collector. The Adams Fourth-Order Variable Step-Size Predictor Corrector Algorithm (12) provided the means of integrating Equations 2 and 3. This technique was chosen because it modifies the step-size when required and needs only two functional evaluations per step while retaining fourth order accuracy. These features provide a quick, yet accurate, solution which is important when this problem is solved on a micro-computer rather than a high speed mainframe computing system. A flow sheet of the scheme employed is provided in Figure 2.

The interception mechanism was included by accounting for the physical size of a real particle. A collision between a particle and a spherical collector occurs when the radial position of the particle is less than or equal to $r_p + r_c$. The dimensionless size of a particle is described by the interception parameter N_R where:

$$N_R = \frac{r_p}{r_c} \quad (19)$$

Collisions on the rear of the collector are neglected since the flow equations do not model the conditions on the downstream side too well. Using the collision criterion that the dimensionless radial position of the particle must be less than $1 + N_R$, the limiting trajectory was determined by first assuming minimum and maximum values of Y_0 and applying the method of bisection to reduce the interval to an acceptable level.

The initial dimensionless starting position, X_s , must be located where the influence of the collector is negligible. Since there is no general agreement in the literature about the selection of this starting position, a trial and error procedure was adopted. This was accomplished by selecting

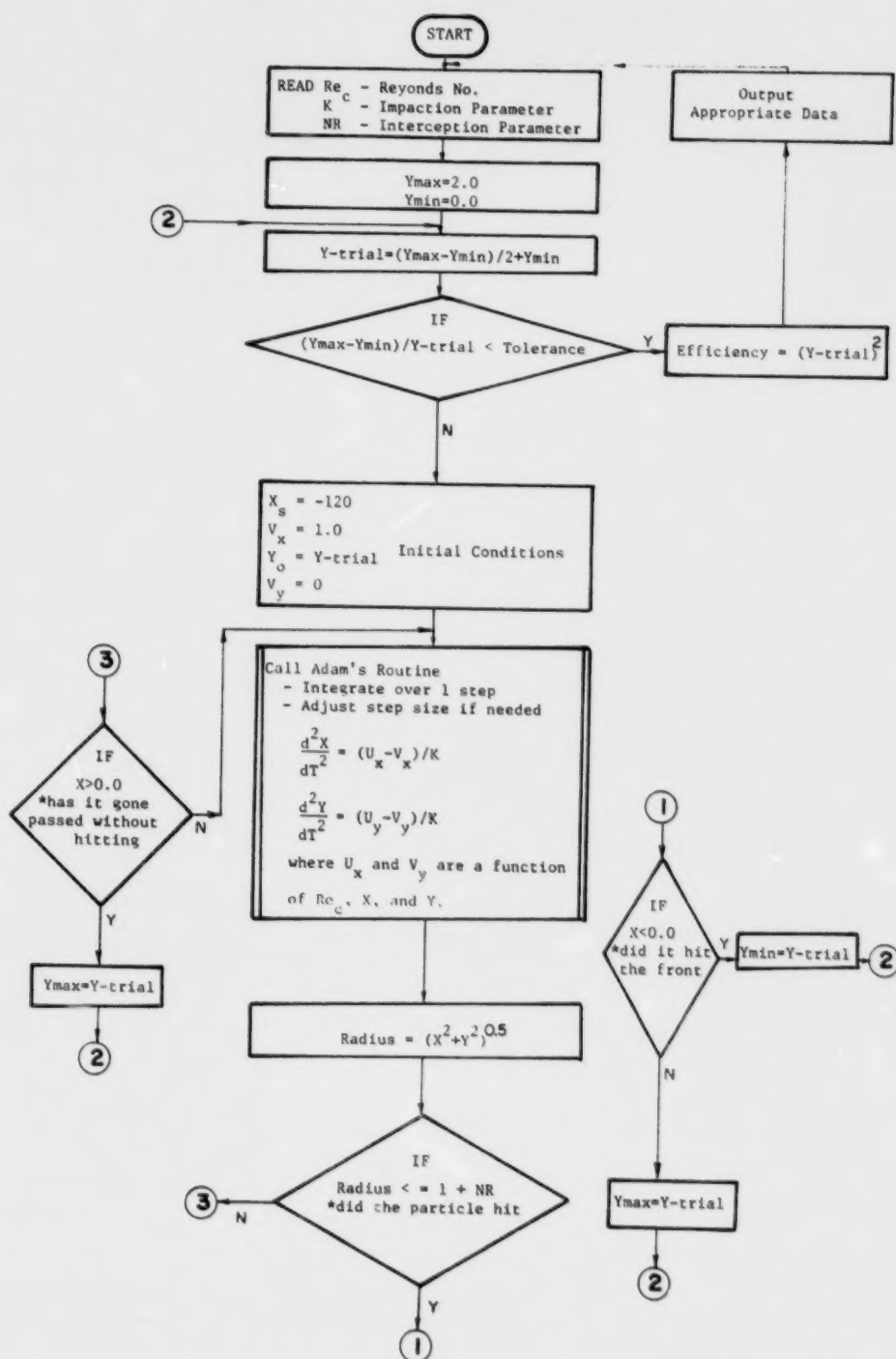


Figure 2. Computational scheme flow sheet.

different values of X_s until no or minimal influence on particle collection efficiency was realized. In this work the value of X_s varied between -75 to -120 for Reynolds numbers of interest.

The validity of the modelling equations and numerical procedures was checked by comparing results to literature values (13) for potential and Stokes flow conditions.

RESULTS AND DISCUSSION

Collection efficiencies have been evaluated for Reynolds numbers ranging from 0 to 80. Figure 3 illustrates the dependence of particle collection efficiency on collector Reynolds number (Re_c) and Stokes number (K) for an interception parameter (N_R) of 0.03. Comparison with the literature was facilitated by selecting $N_R = 0.03$. The values calculated for the limiting cases of potential and Stokes flow are in excellent agreement with the literature (13).

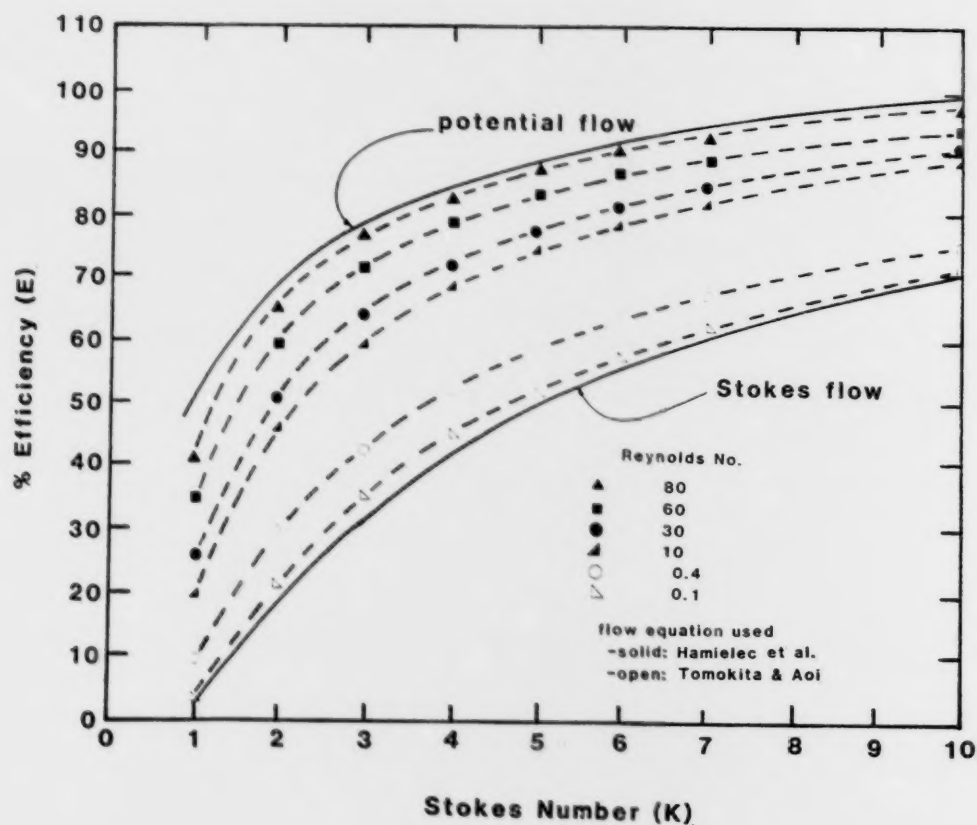


Figure 3. Single droplet collection efficiencies.

It is interesting to note that the collection efficiencies calculated from the Hamielec equation (10) approach the potential flow values when the Reynolds number is about 80. Consequently, for collection efficiency evaluations, this value of Reynolds number appears to be indicative of a flow pattern change from the intermediate to the potential region around the collector.

A critical examination of the results obtained using the Hamielec equation (10) (illustrated in Figure 3) suggests that there must be very significant increases in collection efficiency with increasing Reynolds numbers, for Reynolds numbers less than 10. This speculation is confirmed by the collection efficiencies predicted from the Tomotika and Aoi equation (11) which is valid for Reynolds numbers less than 1.

In order to facilitate application of the improved single drop collection efficiencies to venturi scrubber performance models, it is necessary to develop appropriate analytical expressions that account for the Reynolds number, Re_c , and Stokes number, K . A Gauss-Newton (14) non-linear curve fitting algorithm was used to derive analytical expressions for potential, Stokes and intermediate flow regions.

Potential flow efficiencies were correlated by the relationship:

$$\frac{E_p^{1/2}}{1+N_R} = 1 - \exp(-A K^B) \quad (20)$$

where

$$A = 1.15(1+N_R)^{2.2} \quad (21)$$

$$B = \frac{0.46}{(1+N_R)^{1.6}} \quad (22)$$

Stokes flow efficiencies could not be fitted satisfactorily by a similar function. Consequently, when the expression originally suggested by Mason (15) was modified to the form:

$$\frac{E_s^{1/2}}{1+N_R} = \frac{1}{1 + \frac{0.15 \ln(K) + 1.75}{K - 0.68}} \quad (23)$$

a good fit was obtained.

Interpolation of collection efficiencies for the intermediate flow region was achieved by assuming the relationship:

$$E_I = E_p \phi + E_s (1-\phi) \quad (24)$$

The value of ϕ was obtained, using the Gauss-Newton technique, in the form:

$$\phi = \tanh(0.175 K^{0.234} Re_c^{0.406}) \quad (25)$$

This hyperbolic function was chosen to represent the two extreme flow conditions. When the value of ϕ is equal to zero, Equation 24 reduces to Stokes flow conditions. On the other hand, potential flow conditions are obtained when ϕ becomes unity.

An interpolation scheme that is commonly used, has been attributed to Langmuir (16). It has the form:

$$E_I = \frac{E_s + E_p \left(\frac{Re_c}{60}\right)}{\left(1 + \frac{Re_c}{60}\right)} \quad (26)$$

A comparison of efficiencies obtained using the Langmuir approach and from the present work is shown in Figure 4. The systematic underprediction of collection efficiencies by Langmuir's correlation over the Reynolds numbers of interest can not be explained since the basis for its development has not been discussed adequately in the literature (16).

CONCLUSIONS

In this work, a single drop collection efficiency model was developed for the evaluation of particulate removal in the intermediate flow regime. It was shown that the transition from intermediate to potential flow occurs when the Reynolds number is about 80. Further work must be done to establish collection efficiencies for Reynolds number values less than 10 by using other appropriate numerical techniques. Also, it is recommended that these calculated efficiencies be compared with results determined using flow equations developed by other researchers to ensure the validity of the efficiency data for the intermediate flow regime.

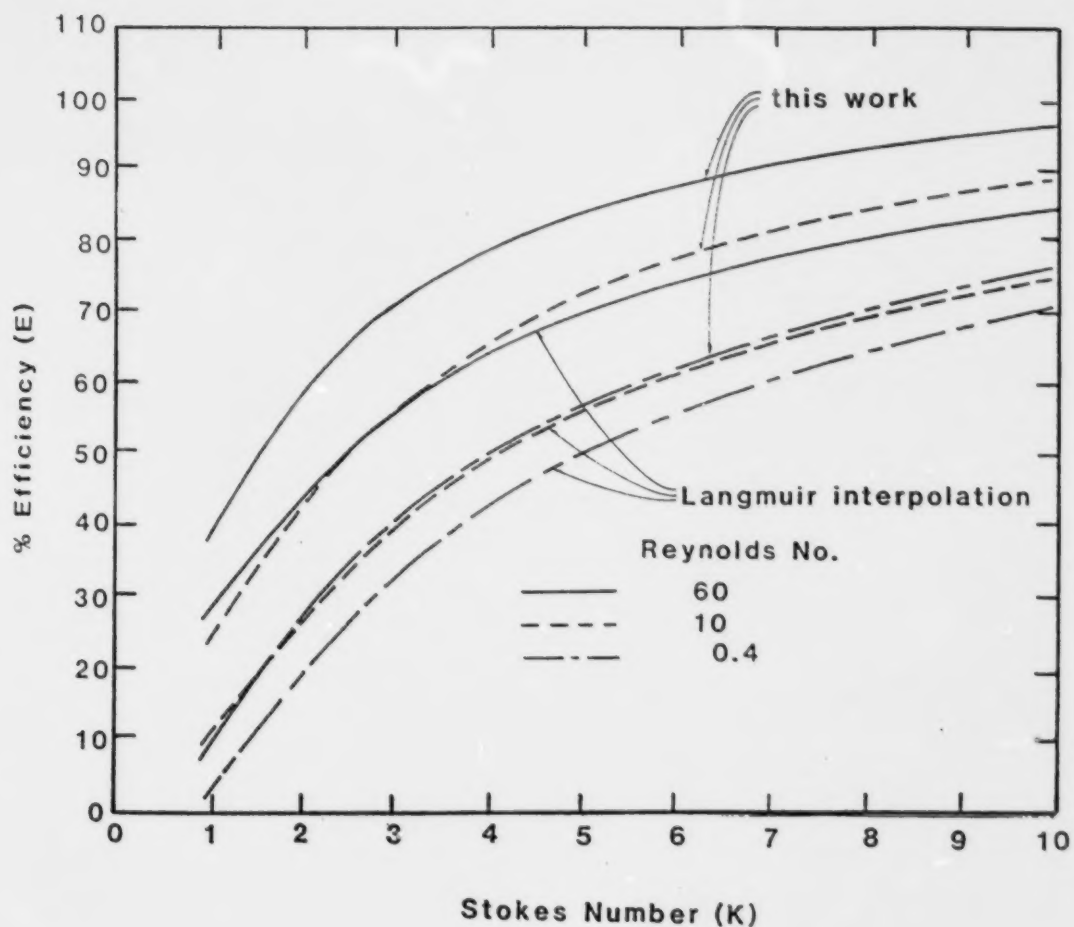


Figure 4. Comparison of interpolation techniques.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

NOMENCLATURE

A	= constant in Equation 20
A_i	= coefficient in the Hamielec equation
B	= constant in Equation 20
B_i	= coefficient in the Hamielec equation
C	= Stokes-Cunningham correction factor
C_D	= drag coefficient
E	= single droplet collection efficiency
E_I	= single droplet collection efficiency for intermediate flow
E_p	= single droplet collection efficiency for potential flow
E_s	= single droplet collection efficiency for Stokes flow
K	= Stokes number, $2 C_o U_o r_p^2 / 9 \mu r_c$
N_R	= interception parameter, r_p / r_c
R	= dimensionless radial coordinate, r / r_c
r	= spherical radial coordinate
Re_c	= collecting drop Reynolds number, $2 \rho_f U_o r_c / \mu$
Re_p	= particle Reynolds number, $2 \rho_f U_o r_p / \mu$
r_c	= radius of collecting droplet
r_p	= radius of particle
T	= dimensionless time, $t U_o / r_c$
t	= time
U_o	= gas stream velocity at an infinite distance from the collecting droplet
U_R	= dimensionless fluid velocity in the r direction, u_r / U_o
u_r	= fluid velocity in the r direction
U_θ	= dimensionless fluid velocity in the θ direction, U_θ / U_o

- u_θ = fluid velocity in the θ direction
- U_x = dimensionless fluid velocity in the x direction, u_x/U_o
- u_x = fluid velocity in the x direction
- U_y = dimensionless fluid velocity in the y direction, u_y/U_o
- u_y = fluid velocity in the y direction
- V_x = dimensionless particle velocity in the x direction, v_x/U_o
- v_x = particle velocity in the x direction
- V_y = dimensionless particle velocity in the y direction, v_y/U_o
- v_y = particle velocity in the y direction
- X = dimensionless coordinate in x direction, x/r_c
- x = x cartesian coordinate
- Y = dimensionless coordinate in y direction, y/r_c
- y = y cartesian coordinate
- y'_o = initial displacement of limiting trajectory for a point mass
- y_o = initial displacement of limiting trajectory for a real particle
- Y'_o = dimensionless initial displacement of limiting trajectory for a point mass, y'_o/r_c
- Y_o = dimensionless initial displacement of limiting trajectory for a real particle, y_o/r_c

GREEK LETTERS

- β = viscosity ratio of dispersed phase to continuous phase
- γ = correction factor for non-Stokesian behavior
- θ = spherical angular coordinate
- μ = fluid viscosity
- ρ_f = fluid density
- ρ_p = particle density
- ϕ = interpolation function defined by Equation 25
- ψ = stream function

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FUME FILTRATION THROUGH A DYNAMIC LIQUID SPRAY CURTAIN

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ABSTRACT

This paper compares the capability of a liquid droplet water spray curtain type impaction scrubber with conventional air pollution abatement devices. Removal efficiency on combustion soot fumes, sub-micron particulate matter, coal and municipal waste flyash, SO_2 and HCl are presented. Innovative new technology for simultaneous $NO_x/SO_x/HCl$ and particulate removal is also discussed.

A rotating disc generates a dense curtain of high velocity liquid droplets. These droplets form a constantly renewed filtration zone through which contaminated air passes. This spray zone behaves like a filter pad formed of fibrous material, but it cannot be plugged. Its gas side pressure drop is below 2 inches water gauge, and its particulate removal and gas absorption capability exceed most high energy venturi scrubbers, but at significantly lower energy inputs. This novel device is named the Rotary AtomizingTM scrubber.

INTRODUCTION

When discussing emissions controls, you've probably heard the comment that "the stack is cleaner than the ambient air". To see how true this is, we have prepared a table regarding particulate emissions, which we will call "How clean is clean?"

TABLE 1. HOW CLEAN IS CLEAN

Particulate Concentration (grains/SCF)	Comment
5.0	uncontrolled boiler stack
0.2	boiler with multiclones
0.02	typical performance standard
<u>0.002</u>	<u>best expectable outlet, controlled</u>
0.0002	unattainable with usual technology
0.0001	primary air standard
0.00003	secondary air standard
0.00001	"good" room air

This is by way of introduction to typical measurements and concepts in air pollution control.

High-speed spinning-disk atomizers are mentioned in Perry's handbook as "primarily useful for spraying slurries and liquids containing solid particles which would clog other nozzles. Hence, they are widely used in spray drying". He goes on to say that a 12-inch disk atomizer spinning at 9,790 rpm would impart the same velocity to a liquid as would a 5,000 psi nozzle, and for the same liquid flow, would require about the same power input. But note, it cannot clog, and continuously pumping the liquid into it requires almost no pressure.

These seemingly unrelated subjects were brought together in the design of the Rotary AtomizingTM scrubber, invented and commercialized by EMCOTEK of Visalia, California during the past five years. Figure 1 presents a single Rotary AtomizingTM unit, or "spool".

DESCRIPTION OF APPARATUS

A typical Rotary Atomizer is driven by an electric motor and gearbox, to atomize the scrubbing liquid into 20 to 40 micron droplets. These are injected radially into an annular scrubbing zone at velocities approaching the speed of sound. This novel high-velocity scrubbing mechanism generates a curtain of water droplets. This liquid droplet curtain behaves like a very fine cloth or paper filter. Removal efficiencies equivalent to cloth filters (baghouses, cartridge filters, etc.) are achieved. The water curtain can not plug or clog, unlike cloth or cartridge filters. There is a constant pressure drop across it, variable only with the gas velocity passing through it. Actual gas side pressure drop ranges between 0.1 to 2.0 inches of water, as presented in Table II.

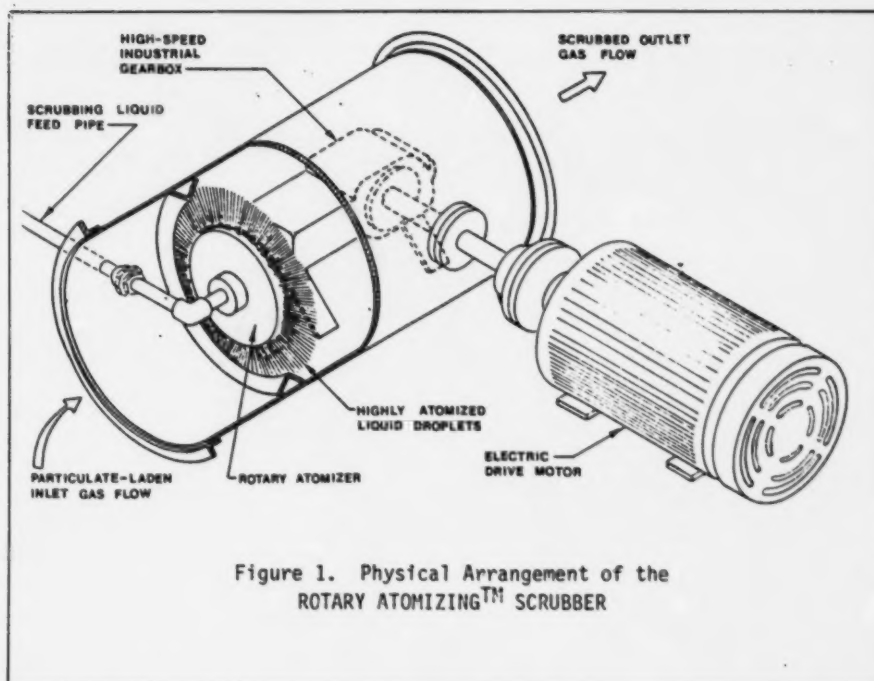


TABLE II. TYPICAL PRESSURE DROP OF GAS
THROUGH SPRAY CURTAIN

Gas Flow (ACFM)	Gas Phase Pressure Drop Through Spray Curtain (inches water column)
1000	0.08
2000	0.85
3000	1.27
4000	1.59
5000	1.82
6000	2.02

It is of interest to note that pressure drop through the spray curtain is independent of particulate loading in the gas stream (1).

The spray chamber has been shaped to conduct the gas through the atomized droplet curtain. Maximum collection efficiency of sub-micron particles is attained by optimizing the distance between the rotating disk and the impaction wall. Optimization is needed both because the water droplet velocity and the density of the water droplet curtain decline toward the periphery of the spray zone, near the impaction wall.

To be most effective, the water leaving the atomizer must remain as liquid droplets, and not flash into steam. For this reason, ahead of the atomizer is a saturation chamber, see Figure 2. It brings the incoming air to 100% relative humidity. (This chamber is also referred to as a pre-spray or quench chamber). Following the atomizer, a demisting chamber and pad is used to collect and separate the water droplets from the gas stream. Demister pads are normally rated for droplet removal of 2 microns and larger. Hence, the demister is not a scrubber, but a water droplet coagulator. EMCOTEK's wire mesh demister is normally designed to operate at between 8 and 16 ft/sec gas velocity. Chevron-type demisters are used in some applications.

EXPERIMENTAL RESULTS

GAS SCRUBBING EFFICIENCY

We have observed that the gas scrubbing efficiency of the Rotary AtomizerTM varies with the flowrate of the contaminated air stream passing through it. At constant spray liquid flow and atomizer RPM, gas outlet contaminant concentration varies as the 4th power of the relative gas velocity through the spray curtain zone. Typical data for SO₂ absorption in pH 7.5 liquid is presented in Table III (1).

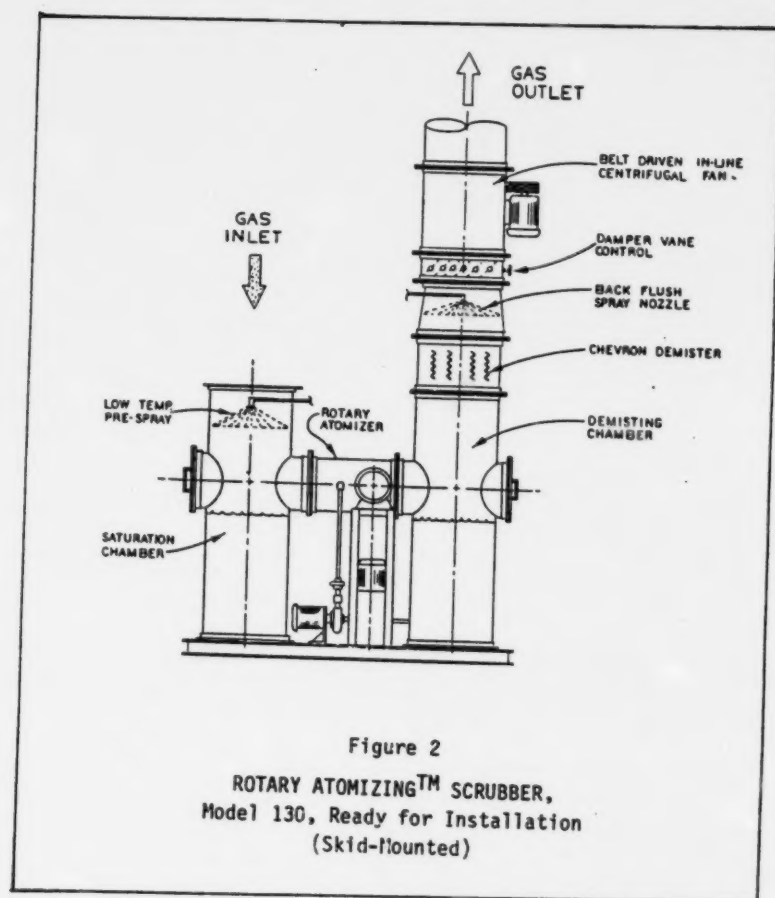


TABLE III. ABSORPTION OF SULFUR DIOXIDE
VERSUS INLET GAS VELOCITY

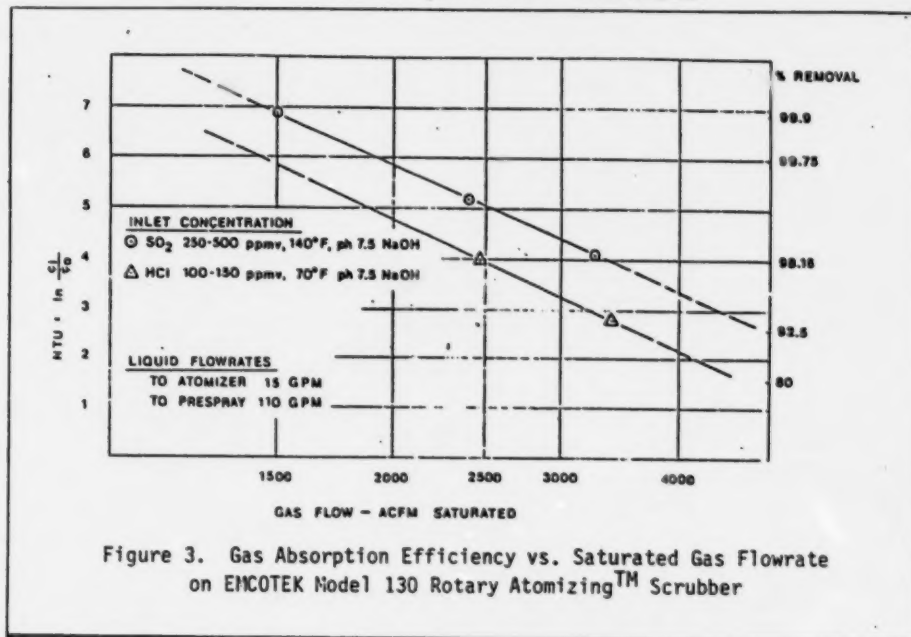
Gas Volume (ACFM)	SO ₂ Absorption Efficiency (Vol %)	Equivalent Number of Transfer Units (NTU)
1500	99.9	6.90
2400	99.45	5.20
3250	98.4	4.13

This observed relationship allows calculation of absorption efficiency from a single experimental run where the inlet and outlet concentrations of the gas and its flowrate have been measured. The equation for this relationship is:

$$\text{Equation [1]} \quad C_i - \left[C_o \left(\frac{V_1}{V_2} \right)^4 \right] \times \frac{100}{C_i} = \text{Vol. \% Removal}$$

Where: C_i = inlet gas concentration (ppmv)
 C_o = outlet gas concentration (ppmv)
 V_1 = new gas flow (ft³/min.)
 V_2 = experimentally measured gas flow (ft³/min.)

This relationship has been plotted for SO₂ and HCl gas absorption in slightly basic (pH 7.5) water, as shown in Figure 3. Equation [1] is quite precise when the prescrubbing and rotary atomizing conditions, such as water spray rate, temperature, inlet gas concentration, and pH of scrubber water are identical between the known and predicted condition.



PARTICULATE SCRUBBING EFFICIENCY

New restrictions proposed by EPA in late 1984 mandate reduction of particulates in the low micron to sub-micron range. EPA's target is changing from total mass emissions to fine particulates, because health officials now believe that it is only the small particles which penetrate deep into the human respiratory system (2).

On particulate removal, we have observed that the outlet particulate concentration varies as the $3/2$ power of the gas velocity through the spray zone. Typical examples are presented in Table IV and plotted in Figure 4:

TABLE IV. SCRUBBING EFFICIENCY OF SUB-MICRON SMOKE
VERSUS INLET GAS VELOCITY

Material	Experimental Grain Loading (grains/SCF)	Gas Volume (ACFM)	Removal Efficiency (%)	
			Calc.	Exper.
TiCl ₄ hydrolysis product (TiO ₂)	0.25	2481	98.5	98.5
	0.079	3408	-	96.6
SnCl ₄ hydrolysis product (SnCl ₄ 5H ₂ O)	0.27	1500	99.6	99.6
	0.27	2500	-	99.2
Diesel Oil	-	1500	93.5	-
Combustion	0.153	2150	90.7	91.0
Soot	0.176	4600	-	71.0

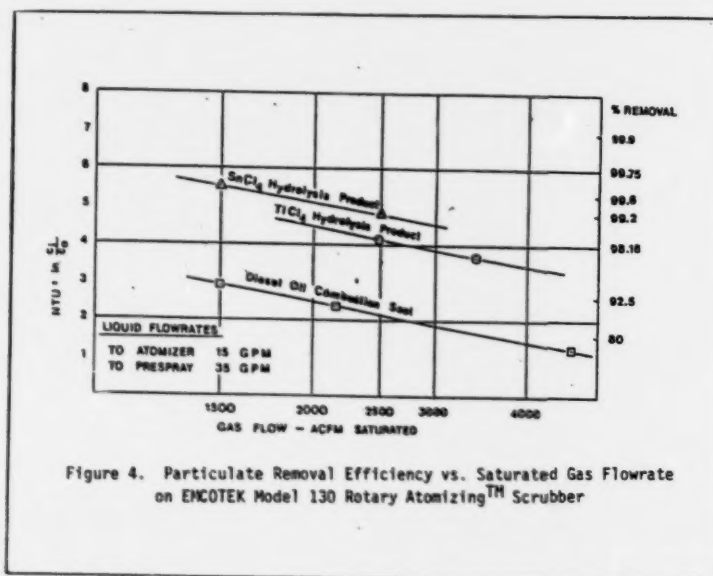


Figure 4. Particulate Removal Efficiency vs. Saturated Gas Flowrate
on ENCOTEK Model 130 Rotary AtomizingTM Scrubber

The equation for particulate removal as a function of gas velocity is presented below:

$$\text{Equation [2]} \quad C_i - [C_o \left(\frac{V_1}{V_2} \right)^{3/2}] \times \frac{100}{C_i} = \text{Wt. \% Removal}$$

Where: C_i = inlet solid loading (grains/SCF)
 C_o = outlet solid loading (grains/SCF)
 V_1 = new gas flowrate (ft³/min.)
 V_2 = experimentally measured gas flow (ft³/min.)

Particle collection in the Rotary AtomizingTM Scrubber results primarily from inertial impaction on the droplets. Efficiency of particulate removal is a complex function of particulate density, inlet size distribution due to prior agglomeration, particle shape, water droplet trajectory, water droplet size distribution, liquid viscosity, and adhesive forces between the water droplet and the particles to be removed. There is an optimum droplet diameter which varies with the rotational speed of the hub and the liquid to gas (L/G) ratio passed through the spray zone. EMCOTEK has taken this optimization into consideration in atomizer design.

SMOKE OPACITY ABATEMENT

Organic Smokes

Wood, oil, and other organic-based smoke and soot particles are generally smaller than 0.9 microns in size, and are very difficult to collect using conventional wet venturi scrubbers or baghouse technology. Baghouses are capable of 98+% removal, but the oil in or on the particles makes them stick to the bag cloth; the bags plug up and cannot be cleaned unless large amounts of dry additive are used to precoat the bag material. There are no baghouse installations which are currently operating on oil-based smoke without exceptionally high operating and maintenance costs. Electrostatic precipitators suffer from similar problems and are not economical on first cost in sizes smaller than about 50,000 cfm.

Afterburners can be used, but only at very high operating cost, since the entire volume of exhaust must be heated to over 1400°F to destroy a miniscule percentage of particles. Catalytic afterburners are another possibility, but at a capital investment which approaches the electrostatic precipitator. Catalysts are also rapidly poisoned by many unsuspected impurities in the gas to be scrubbed.

Prior to the invention of the Rotary AtomizingTM scrubber, no one in the USA could successfully scrub combustion soot with water. The hypothesis was that water, since it is polar, will not remove nonpolar soot. But unlike the venturi scrubber, which converts mechanical to gas energy, and then into water breakup, the Rotary AtomizingTM scrubber puts mechanical power directly into generation of a huge number of tiny, high velocity water droplets. A curtain of these high velocity liquid droplets removes smoke and soot pollution. Typical opacity reductions on organic smokes are presented in Table V.

TABLE V. TYPICAL OPACITY REDUCTIONS ON ORGANIC SMOKES

Smoke Abatement Applications	Inlet Opacity	Outlet Opacity
Combustion (oil) soot-typical	33	9
Oil Smoke (from wire tempering bath)	40	8
Magie Oil (from offset printing press)	37	1
Di-Octyl Phthalate Fumes	41	1.2
Smoking Hardwood (smokehouse)	15	2
Castor Oil Smoke (fiberglass manufacture)	35	5
Diethylene Glycol Fumes	39	<1
Butyl Tin Chloride Fumes	22	3.5

Inorganic Smokes

Opacity reductions achieved on the Rotary Atomizer with inorganic smoke particulates are presented in Table VI.

TABLE VI. TYPICAL OPACITY REDUCTIONS ON INORGANIC SMOKES

Type Smoke	Inlet Opacity	Outlet Opacity
Titanium Oxide Smoke from $TiCl_4$ Hydrolysis with moist air.	30	2.5
	98	18
Silicon Tetrachloride hydrolysis with moist air.	48	10
Tin Tetrachloride hydrolysis with moist air.	23	0.5

FLYASH REMOVAL

Flyash emission using a Rotary AtomizingTM scrubber were tested at ambient temperature and gas flowrates between 3000 and 6000 ACFM. Two types of flyash were available for emissions testing. The first was coal ash (from an electrostatic precipitator) and the second a municipal waste flyash from Lassen Jr. College (baghouse). Each of these flyash samples was fed at 0.2 grains/SCF to a jetmill and then introduced to the bulk gas stream.

The demister used consisted of a Yorkmesh pad. The air flow was vertical through this demister. Its recommended design velocity is 10-15 ft/sec. At 3000 ACFM, the actual velocity was 8.4 ft/sec; and at 6000, it was 16.8 ft/sec. Hence, the design range was undershot and overshot on the 3000 and 6000 ACFM tests, respectively. The flyash dust emissions observed on all tests were substantially below 0.01 grains/SCF, the target needed to meet California's air pollution requirements for solid waste-to-energy plants.

The particle size distribution and loading for the Lassen Jr. College flyash was measured at two locations. The first location was the inlet to the prespray, and the second was at the stack's outlet. This data shows the fraction removed by the prespray, and the fraction of material which was actually scrubbed by the rotary atomizer. See Table VII for results from these tests.

TABLE VII. CUMULATIVE AND STAGE-WISE
REMOVAL EFFICIENCY OF FLYASH.

Type Flyash and Stage Eff.	Dust Loading (grains/SCF)		Air Volume	D ₅₀	Geom. Distrib.	Stage Efficiency	Cumulative Efficiency
	Inlet	Outlet	CFM	(μm)	(σ _g)	(%)	(%)
<u>Municipal Waste Ash</u> (Lassen Jr. College Baghouse)							
Prespray	0.20	0.040	3000	5.1	2.35	80	80
Rot. Atomizer	0.040	0.0010	3000	2.4	2.33	97.4	99.45
Across Scrubber	0.20	0.0012	6000	5.1*	2.35	-	99.4
<u>Coal Boiler Ash</u> (ESP)							
Across Scrubber	0.20	0.00083	6000	3.8§	2.16	-	99.6
* 20% less than 2.4 microns							
§ 20% less than 2.2 microns							

Table VII shows that 20% of the municipal waste flyash has a D₅₀ of 2.4 microns, the more difficult size range for removal. The rotary atomizer removed 97.4% of this fraction at a demister efficiency below optimum. Optimization of the demister would raise the removal efficiency of this stage above 97.8%. In the first test above, 80% was removed by the prespray tower. Hence, the rotary atomizer acted as a polishing step for particles in the 0.3 to 2.4 micron cut size range. The overall ash removal efficiency was 99.45% for the total system.

Flyash from a coal boiler was removed even more readily by the EMCO-TEK scrubber at 6000 ACFM. This was probably due to the higher specific gravity of the coal flyash.

The D_{50} particle sizes observed on the jet-milled flyash are in the range reported by Calvert and Englund for Municipal Waste Flyash (Handbook of Air Pollution Technology - 1984).

CHOOSING A SCRUBBER

As a first step in choosing a scrubber, one needs to know the grain loading, particle size distribution, and properties of the particles to be removed. This data can be obtained with a Cascade Impactor test typified by Figure 5 data. The Impactor test provides the particle size distribution, and D_{50} (the size which 50% of the particle weight is smaller than), and the geometric standard deviation, σ_g , another measure of the distribution (3).

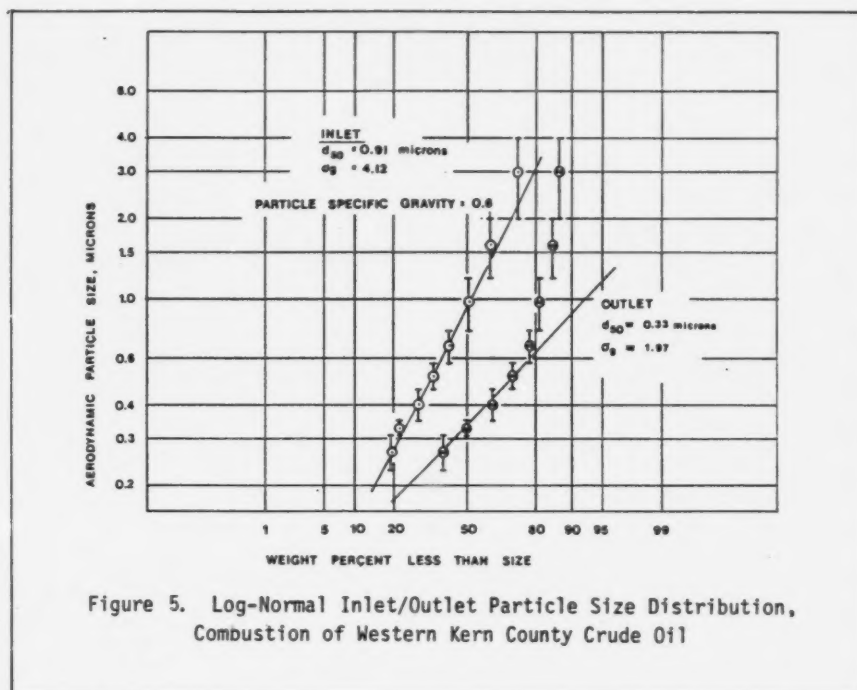
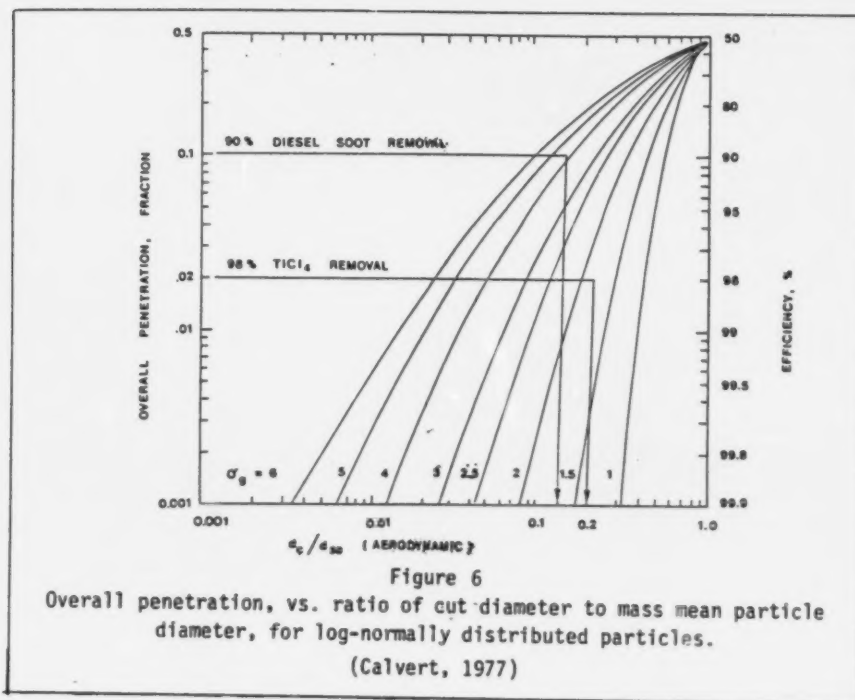
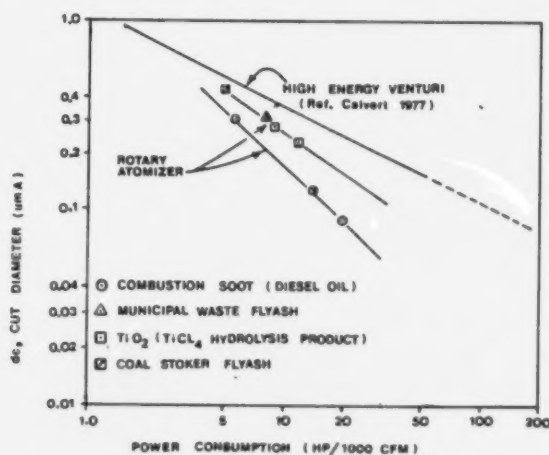


Figure 5. Log-Normal Inlet/Outlet Particle Size Distribution, Combustion of Western Kern County Crude Oil

With these measurements, a cut diameter, d_c , is chosen which meets the governing efficiency requirements. Calvert's plot for calculating the cut diameter is presented as Figure 6.



Over the years, in papers by Calvert (4) and Semrau (5) the effectiveness of different types of scrubbers has been correlated against power consumption using these parameters. Since Calvert's correlation is the simplest, we have presented his correlation in Figure 7 in which a comparison on power input is made between a high energy venturi scrubber and the Rotary Atomizing™ Scrubber.



For example, Figure 7 shows that if a cut diameter, d_c , of 0.2 microns (Aerodynamic) is needed, one could choose either a Rotary AtomizingTM scrubber requiring 9 to 14 HP/1000 ft³/min. or a high energy venturi scrubber requiring 30 HP/1000 ft³/min. If a combination of particulate and gases must be removed simultaneously, such as SO₂ plus soot from an industrial boiler, then the options are limited to an ESP or baghouse followed by a packed tower, a high energy venturi, or a Rotary AtomizingTM Scrubber.

Although initial selection of scrubber type and size can be calculated, a pilot scale test on a side stream of an existing plant process is the most reliable way to confirm the design. EMCOTEK has a small, portable scrubber unit available for such tests at its client's plants.

DEVELOPMENT OF NEW EMISSION ABATEMENT TECHNOLOGIES BY EMCOTEK

ABATEMENT OF COMBINED NO_x/SO_x/HCl/PARTICULATE EMISSIONS

In combination with BTU Services, Inc. of Visalia, EMCOTEK has demonstrated a new process to remove 70-90% NO_x in combination with flyash, SO_x and HCl. This process is applicable to municipal solid waste incineration, and crude oil, biomass, or coal-fired boiler applications. Removal of combined emissions is compared with central California requirements in Table VIII.

TABLE VIII. COMBINED NO_x/SO_x/HCl/PARTICULATE EMISSION

Pollutant	Central California Requirements (w/out offsets in attainment areas)	Test Results Using EMCOTEK- BTU Processes
Particulate Emission (grains/SCF)	0.01	0.002
Sulfur Oxides (ppm)	<72*	<5
Nitrogen Oxides (lbs./day)	250	25-75§
Hydrogen Chloride (ppm)	-	<2

* Kern County, California standard, or 95% removal.

§ Based on a 70-90% reduction from inlet level of 360 ppm NO_x, 3% O₂, dry when firing crude oil at 7MM BTU/hr.

In the BTU Process (patents pending), an initiator is injected into the low temperature zone of the steam boiler. This initiator oxidizes NO to NO₂. The initiator material is commercially available and does not involve ammonia or urea. The oxidation reaction is a non-catalytic vapor-phase reaction. When the technologies are combined, the EMCOTEK-BTU Processes create NO-depleted boiler off-gases, which pass through a hot cyclone to remove large flyash particles and are then quenched in the prespray section of the scrubber. The prespray also removes most particles above 2.5µm. The Rotary Atomizer™ spray curtain is then used to remove the remaining fine particles.

Inexpensive inorganic reagents are continuously added to recirculated scrubber liquid, which comprises the spray curtain. These reagents remove SO₂ and HCl.

The atomizer hub is followed by a scrubbing tower through which scrubber liquor is also recirculated. NO₂ remaining in the boiler off-gas is decomposed to molecular nitrogen gas in the scrubber tower. The overall NO_x removal efficiencies during crude oil-firing tests were measured at the scrubber stack. Without the initiator material injected into the boiler, the NO_x emission at the scrubber stack was 360 ppm at 3% O₂, dry. When the initiator was injected, the NO_x emission at the scrubber stack was reduced to as low as 44 ppm at 3% O₂, dry. The variation in NO_x removal with amount of initiator material is presented in Figure 8.

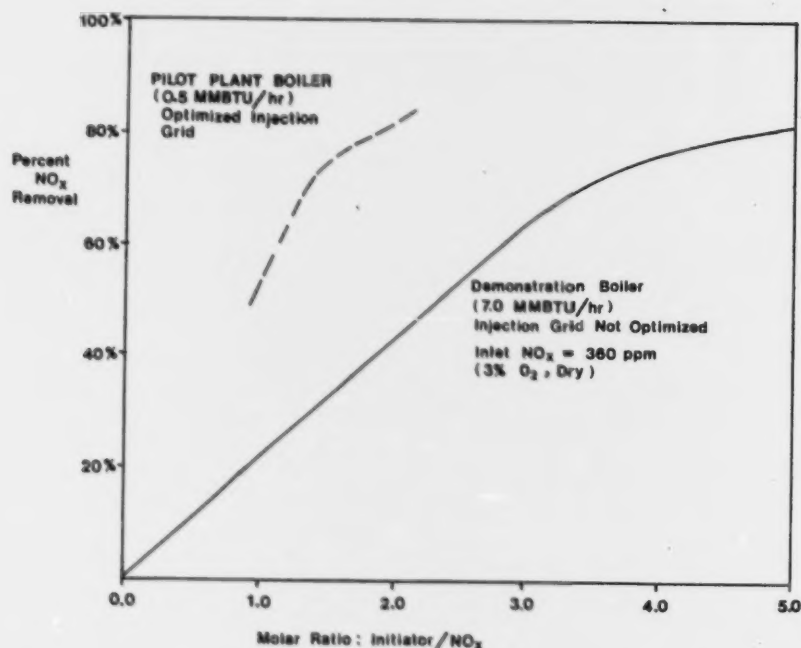


Figure 8. Effect of Quantity of Initiator Material on NO_x Removal in a Caustic Scrubber

As seen in Figure 8, to achieve 75% NO_x removal from an inlet NO_x level of 360 ppm requires an initiator to NO_x molar ratio of 4.0. In this case, the initiator material would cost about \$650/ton of NO_x removed. Further optimization of the injection system should allow similar performance at a lower cost.

In comparison with using ammonia or urea for boiler injection to accomplish NO_x control, this new process may provide superior NO_x removal without objectionable ammonium salt emissions and air heater fouling associated with ammonia or urea technologies. Demonstration of this new technology for combined emission control on larger-scale commercial equipment is anticipated in the near future.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

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Session C-2: OPACITY/MEASUREMENTS

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OPTIONS TO ALLEVIATE PLUME OPACITY PROBLEMS
CAUSED BY CONDENSATION AEROSOLS

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ABSTRACT

The opacity of "detached plumes" formed by condensation of vapors depends upon both the concentration of condensible vapors and the in-stack concentration of fine, submicron, particulate matter. This paper provides an analysis of the condensing aerosol problem and an evaluation of possible control approaches to reduce the downwind detached plume opacity. The opacity of such plumes may be reduced by reducing the concentration of condensible vapors or the in-stack concentration of fine particles or both. The results of the analysis indicate that for low concentrations of condensible vapors the detached plume opacity may be adequately controlled by reducing the in-stack fine particulate concentration alone. For high concentrations of condensible vapors, however, reduction of in-stack fine particulate concentration alone may not be effective, and reduction of vapor concentration may be necessary along with particulate removal for adequate reduction of plume opacity. Different combinations of levels of reduction of vapor concentration and particulate phase concentration are possible to achieve a desired result; and thus may be optimized to obtain a cost-effective combination.

This paper has been reviewed in accordance with the U.S. Environmental Protection Agency's peer and administration review policies and approved for presentation and publication.

INTRODUCTION

BACKGROUND

In a smoke plume resulting from a combustion process, the opacity may increase in the region of the plume where hot stack gas is mixed with the cooler ambient atmosphere. The phenomenon is often called a "detached plume" because the plume is not continuous at the point of emission. In a broad sense, the plume exhibits an increase in downwind opacity as it is mixed with ambient air. Interest in a detached plume is often aroused because of regulatory issues. The fact that the opacity of such a plume at the stack is lower than the downwind opacity indicates that the particulate phase at the stack has lower optical obscuration than the downwind plume. The downwind plume opacity thus cannot be related to the particulate mass concentration and size distribution at the stack in a simple manner. Often the particulate mass concentration at the stack as measured by conventional stack tests is well within compliance limits, but the downwind plume opacity is in violation of opacity regulations.

The increase in plume opacity is due to formation of additional optically effective material within the plume. Condensation of vapors such as sulfuric acid vapors, as the plume is cooled by ambient air, is the dominant reason in most cases. Generally, the opacity of a detached plume increases fairly quickly, within 1-2 seconds, after the plume is released from the stack. This indicates that the condensing vapors are probably already present in the stack gases rather than being formed by any secondary reaction mechanisms.

The problem posed by such primary condensible vapors has only recently been recognized, and a systematic strategy or technology to control the visible emissions resulting from materials condensing downwind in the plume has not yet been developed. This paper provides a better understanding of the condensing aerosol problem and an evaluation of possible control approaches to reduce downwind detached plume opacities.

LITERATURE SURVEY

Most related information on detached plumes either is unpublished or has not been developed in a systematic manner. Conner (1976) reported a significant difference between in-stack opacity at a utility boiler firing high sulfur oil and plume opacity measured downwind with a plume LIDAR (light detection and ranging technique). Opacities over 50% were observed in the plume which were believed to be due to condensed sulfuric acid. Probably, the first reported observations of the detached plume opacity, directly relating it to the measured sulfuric acid vapor concentrations,

were by Nader and Conner (1978) and were presented at a workshop on primary sulfate emission from combustion sources. A number of cases, where the in-stack particulate matter is in compliance with the established standards, yet the downwind opacity exceeds the opacity standard, are currently under negotiation between EPA and industrial and utility companies.

A number of studies have been made to determine the amount of sulfuric acid vapor present in power plant stack gases and to relate it to process operating conditions such as percent excess air used, percent of sulfur in fuel, concentration of SO_2 in the combustion gases, and presence of vanadium which acts as a catalyst for SO_3 formation from SO_2 (e.g., Homolya and Cheney, 1978; Hunter and Engle, 1978; and Dietz et al., 1978). Some of the studies have focused on combustion modifications to reduce the amount of acid vapors present (Levy et al., 1978). In general, combustion of low sulfur fuel and low excess air/oxygen produce lower sulfuric acid vapor concentrations. The problem of detached plume opacities tends to be more severe for combustors firing oil, which more frequently contains high vanadium content, than those firing coal. Also, oil combustion produces less particulate matter (to obscure the effect of vapors) than coal combustion, and the particulate phase is generally high in submicron particle content.

Recently, a mathematical model was developed by Damle et al. (1984) to quantitatively determine the effect of condensing vapors on the plume opacity. This model includes plume dynamics and the vapor condensation and aerosol processes occurring within the plume as the plume travels downwind. The model indicated a strong influence of both the condensible vapor concentration and the concentration of submicron particulate matter already present at the stack conditions on the peak opacity of the detached plume. These indications were confirmed by another parallel study by Yung et al. (1984) who experimentally studied sulfuric acid vapor condensation under different process conditions. There are only a few actual field measurements of the "detached plume" opacities, and the measurements often lack detailed information regarding in-stack particle size distribution and concentration, water and sulfuric acid vapor concentrations, and ambient meteorological conditions. However, those measurements and the predictions by the plume model with assumed reasonable values for missing parameters do relate the condensation of vapors to the detached plume opacities.

DESCRIPTION OF DETACHED PLUME PHENOMENON

In the case of a "detached plume," the plume usually has very little or no visible emissions at the stack. The opacity may increase suddenly or gradually as the plume travels downwind. In a broad sense, a detached plume may be considered a plume that exhibits higher downwind opacity as compared to the in-stack opacity. Condensation of materials present as vapors at the in-stack conditions is considered to be the dominant cause of the detached plume opacity problem due to the short time after plume release from the stack in which the peak plume opacity appears. The following section briefly describes the condensation and aerosol growth processes occurring in the plume as it travels downwind. This also was the basis of the mathematical model developed by Damle et al. (1984). The formation and growth of aerosols

affect the particulate concentration and size distribution and thus affect the plume opacity.

PLUME DYNAMICS

After emission from a stack, a plume rises due to the initial momentum and buoyancy of the hot plume gas. As the plume rises, it entrains cooler ambient air that cools and dilutes the plume. The rate of entrainment or mixing of the ambient air in the plume primarily depends on plume buoyancy and the velocity differential between the plume and wind. This rate depends on the time elapsed after release of the plume from the stack. Initially, the dilution rate is, of course, maximum: it decreases as the plume is diluted progressively by the ambient air. The principal parameters affecting the rate of plume dilution and cooling are the plume velocity and temperature, ambient temperature and wind speed, and the atmospheric turbulence and the lapse rate. To predict condensation of vapors and to determine its rate, it is necessary to follow the vapor concentration and temperature variations with time after plume release. By setting up appropriate mass, momentum, energy, and species balance equations across a plume section and following this plume section with time, the change in the plume properties with time is determined (Damle et al., 1984).

VAPOR CONDENSATION

As the plume cools, the plume temperature may drop below the dew point for the condensable vapors, thus commencing condensation. Figure 1 is a schematic of the condensation and aerosol processes occurring in a plume. Two primary approaches are available for the condensation process -- heterogeneous condensation and homogeneous nucleation. The relative contribution of the two mechanisms to the condensation process at a given time depends upon the supersaturation ratio of the condensable species, which is the ratio of the partial pressure of the vapor to its equilibrium vapor pressure at the prevailing conditions.

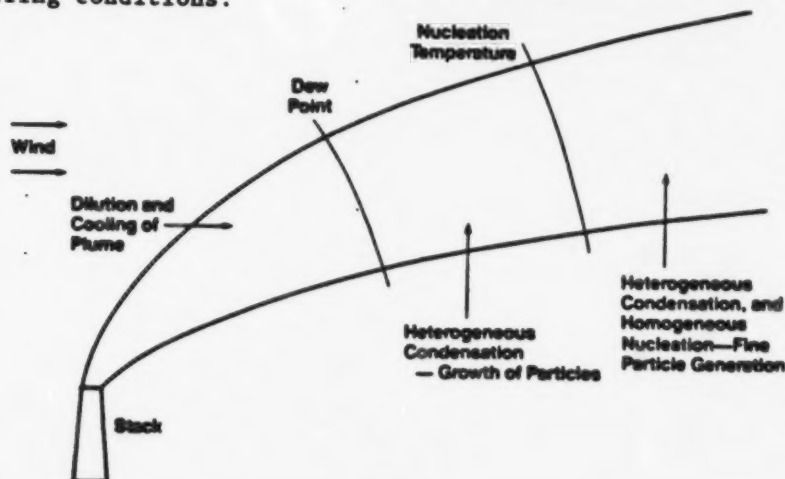


Figure 1. Schematic of condensation and aerosol processes occurring in a plume.

- 1) Heterogeneous condensation. Upon reaching the dew point, vapors may begin to condense heterogeneously on the available surface area of existing particulates in the plume. For low supersaturation ratios (< 5), this provides the dominant condensation mechanism. The rate of heterogeneous condensation on a particle is given by the rate of diffusional transport of vapors to the particle surface and depends on the vapor concentration gradient and the diameter of the particle. Relations to determine the rate of heterogeneous condensation and corresponding growth of particles are well developed for particles in both continuum and molecular regime (Friedlander, 1977). An approximate relation applicable to the entire particle size range has also been proposed by Fuchs and Sutugin (1971).
- 2) Homogeneous nucleation. As the vapors condense heterogeneously, the supersaturation ratio of the condensible species may increase or decrease depending on the rate of heterogeneous condensation of the vapors and the rate of continued cooling and dilution of the plume by ambient air. If sufficient surface area is not available for the heterogeneous condensation process, the supersaturation may increase to a point where the homogeneous nucleation of the vapors becomes appreciable. The temperature at which the nucleation process becomes significant, generating at least 1 droplet/sec/cm³, is termed "nucleation temperature." The homogeneous nucleation process generates many very fine droplets, of a stable critical size, which in turn provide a large surface area for the condensation to proceed by the heterogeneous mechanism. The actual amount of vapor condensed by this mechanism is often very small but it provides a large surface area on which the condensation can proceed heterogeneously. The critical size of the generated droplets depends on the supersaturation ratio and the physical properties of the condensed phase. Classical nucleation theory may be used to determine the stable critical size and the rate of generation of droplets of this critical size (Friedlander, 1977).

GROWTH OF PARTICLES AND PLUME OPACITY

Growth of existing particles by heterogeneous condensation and formation of new particles by homogeneous nucleation change the particulate size distribution and concentration in the plume. In addition, interparticle collisions also change the size spectrum of particles, although this is significant only for very fine particles. The change in the concentration and size distribution of the particulate phase in the plume determines the change in the plume opacity. The opacity of a plume may be determined by (Conner and Hodgkinson, 1972):

$$O_p = 100 [1 - \exp(-2Rb_{ext})] \quad (1)$$

where O_p is the percent plume opacity, R is effective plume radius, and b_{ext} (cm⁻¹) is overall extinction coefficient of the particles, and is given

by:

$$b_{\text{ext}} = \pi/6 \int_{d_{\text{min}}}^{d_{\text{max}}} Q_i d_i^3 n_i dd_i \quad (2)$$

where Q_i is specific extinction efficiency per unit mass for a particle of size d_i , and n_i is the number concentration of particles of size d_i .

The increase in the plume opacity is due to formation of additional optically effective material within the plume. The particles in the 0.2-1.0 μm diameter range are particularly effective in obscuring visible light on a unit mass basis, with a peak obscuration at 0.5 μm particle diameter (Friedlander, 1977). Figure 2 shows scattering efficiency per gram, Q , as a function of particle diameter for a particle refractive index of 1.5-0.05i and a light wavelength of 550 nm. The ultrafine particles (less than 0.1 μm diameter) are actually less effective than the above size range, but have a potential to grow into the optically most effective size range by the condensation process. The fine particles also offer a large surface area per unit mass for the condensation process. Thus, particles smaller than 0.1 μm diameter grow into the optically most effective size range by the heterogeneous condensation process. The submicron particles present in the plume are thus very important in determining the plume opacity for two reasons: 1) particles already in the optically effective size range, and 2) smaller particles which can rapidly grow into that range.

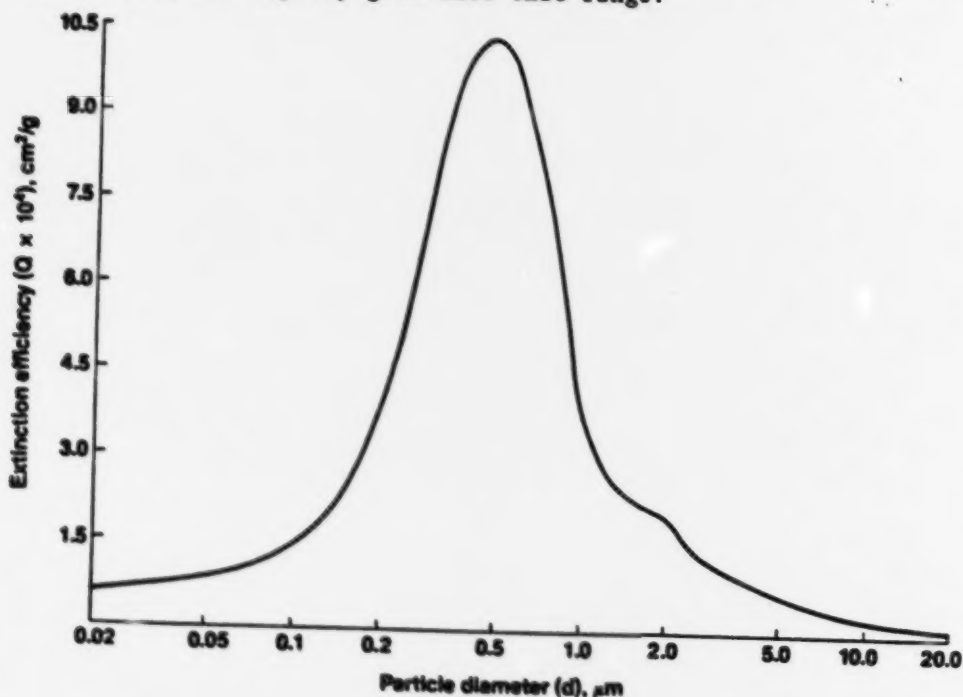


Figure 2. Specific extinction efficiency per unit mass of particles.

The characteristic particles in the 0.1-0.15 μm size range, typical of the coal fired power plant fly ash size distribution (Flagan and Friedlander, 1978; Damle et al., 1982), thus play a very important role in determining the plume opacity. The measurements by Markowski et al. (1980) indicated the presence of such a submicron mode at both the inlet and outlet of a high efficiency ESP following a pulverized-coal-fired utility boiler. Measurements at six coal fired boilers by McElroy et al. (1982) also identified such distribution. Submicron particles are even more significant for oil fired combustors: they can constitute almost half of the total particulate emissions, as indicated by measurements of Mormile et al. (1982). The submicron particles appear to be a general feature of coal or oil combustion and are produced by a vaporization/condensation mechanism. In the absence of any fine particles in the plume, the homogeneous nucleation process becomes important and generates many ultrafine particles which may eventually grow by condensation and coagulation into the optically effective size range. In such a case, the rate of growth of the ultrafine particles with respect to plume dilution becomes important in determining the plume opacity. For high vapor concentrations, the growth rate may become high enough for the particles to grow into the optically effective size range, thereby dramatically increasing the plume opacity.

PARAMETERS CONTROLLING THE CONDENSATION PROCESS

The condensation process as described above depends on a number of parameters that determine the condensation mechanism and the rate of the condensation. The primary variables are the condensable vapor concentrations and the concentration and size distribution of the particulate phase, especially of the submicron fraction present in the plume. Parameters affecting the entrainment or dilution rate of the plume are also important: they determine the change in the plume velocity, temperature, and vapor and particulate concentrations with the plume's travel downwind. These parameters include plume velocity, ambient wind speed, plume buoyancy with respect to ambient air, and atmospheric lapse rate. The rate of plume dilution and cooling also determines the time available for the heterogeneous condensation process before the homogeneous nucleation becomes significant. The effect of these various parameters on the plume opacity is discussed in the following sections.

PLUME OPACITY PREDICTION

PLUME MODEL

To determine the opacity of the plume as a function of plume travel, the particle size distributions and concentrations need to be obtained, along with the effective plume radius, as a function of time elapsed after plume release from the stack. The mathematical model developed by Damle et al. (1984) is dynamic in nature and follows a plume as it travels downwind. The conservation equations of mass, momentum, energy, and species mass are solved to determine the plume properties such as velocity, effective radius, temperature, and species concentration as a function of time or the distance traveled from the stack. By using the gas-phase vapor species concentra-

tion, a dew point is determined. If the gas-phase temperature is lower than the dew point, the rate of vapor condensation, growth of existing particles, and the rate-of formation of new particles are determined by appropriate equations for both condensation mechanisms. Using the updated particulate concentration and size distributions, plume opacity is determined as a function of time.

The computer program "PLUME" was written in FORTRAN to solve all the pertinent equations and simulate a plume. The details of the program are reported elsewhere (Damle et al., 1984; Owen and Viner, 1985). Plume simulations were carried out using this program to determine the effect of various parameters on the plume opacity as discussed below.

MODEL SIMULATIONS

The most frequently encountered condensible material in detached plumes is sulfuric acid. Sulfuric acid is also very hygroscopic, and its equilibrium properties are strongly related to the water vapor content of a plume. Therefore, binary condensation of sulfuric acid and water is considered in the model simulations described here. Power plant fly ash emissions from both coal and oil fired boilers typically exhibit a bimodal particle size distribution, with fine particles of about $0.15\text{ }\mu\text{m}$ diameter. The mean size of the large particles is difficult to determine, especially for the uncontrolled emissions, since the first stage in the cascade impactor usually collects particles larger than $10\text{ }\mu\text{m}$. For controlled sources, the large particles may be $4\text{--}6\text{ }\mu\text{m}$ in diameter. For the simulations presented here a bimodal lognormal size distribution was used: the mean size of fine particles was $0.15\text{ }\mu\text{m}$ diameter, and that of large particles was $4\text{ }\mu\text{m}$. The proportions of the respective modes depend on the type of fuel used and the efficiency of the particulate control equipment. For the base case simulation, 10:1 ratio was used with a large particle concentration of 100 mg/m^3 and that of fine particles of 10 mg/m^3 .

To assess the effect of various parameters identified in the previous section, several plume simulations were carried out. A base case was first established by using typical ambient and stack gas conditions. A sulfuric acid concentration of 10 ppm and water vapor content of 5% by volume in the flue gas were used in the base case. A single component was then varied with respect to the common base case to determine the effect of that variable. The following base case parameters were used in these simulations:

Initial plume parameters (at stack exit):

- Diameter, 4 m
- Gas temperature, 200°C
- Velocity, 20 m/s
- Inclination, 90 degrees (vertical)
- Water vapor concentration, 5% by volume
- Sulfuric acid vapor concentration, 10 ppm
- Large particle concentration, 100 mg/m^3
 - number mean diameter, $4\text{ }\mu\text{m}$
 - geometric standard deviation, 1.8

Fine particle concentration, 10 mg/m³
 number mean diameter, 0.15 μ m
 geometric standard deviation, 1.3
 Molecular weight of gases, 29

Ambient Parameters

Wind speed, 5 m/s
 Temperature, 20° C
 Relative humidity, 50%
 Sulfuric acid concentration, 0
 Particulate concentration, 0

The results of these simulations are shown in Table 1 which shows that the most influential parameters are the in-stack concentrations of both sulfuric acid vapor and fine particles. The ambient parameters wind speed and the corresponding entrainment rate do have a moderate influence on the plume opacity; however, they are not controllable parameters, and obviously cannot be considered in the control approaches.

TABLE 1. EFFECT OF PARAMETERS ON PEAK PLUME OPACITY
 (Simulation Study)

No.	Case	Initial Opacity %	Peak Opacity %
1	Base	11.1	38.1
2	Low Sulfuric Acid Vapor Concentration (3 ppm)	11.1	18.8
3	High Sulfuric Acid Vapor Concentration (30 ppm)	11.1	73.9
4	Low Fine Particle Concentration (1 mg/m ³)	5.1	13.8
5	High Fine Particle Concentration (100 mg/m ³)	53.8	68.8
6	Low Large Particle Concentration (10 mg/m ³)	7.4	35.7
7	Low Water Vapor Concentration (1%)	11.1	37.5
8	High Water Vapor Concentration (20%)	11.1	34.3
9	Low Initial Plume Temperature (125°C)	11.1	39.2
10	Low Wind Speed (1 m/s)	11.1	41.8
11	High Wind Speed (20 m/s)	11.1	19.3
12	Low Ambient Relative Humidity (5%)	11.1	38.2
13	High Ambient Relative Humidity (95%)	11.1	37.2
14	Low Ambient Temperature (1°C)	11.1	35.0

To further illustrate the effect of in-stack concentrations of sulfuric acid vapor and the fine particle fraction of total particulates, additional simulations were carried out: the fine particle concentration was reduced to 2 mg/m^3 (about 2% of total particulate mass), and the mean size for the large particle mode was lowered to $2 \text{ }\mu\text{m}$. The results are shown graphically in Figure 3. In spite of the lower concentration of the fine particulate fraction, its presence has substantial effect on the plume opacity as seen by the difference between the peak opacity with the bimodal distribution and that with only the large particle fraction. The reasons for the effectiveness of the fine particle mode have already been discussed. Note that the mean size of the submicron fraction is also important in determining the plume opacity and that somewhat different results will be obtained by changing the mean size and concentration of the submicron size fraction.

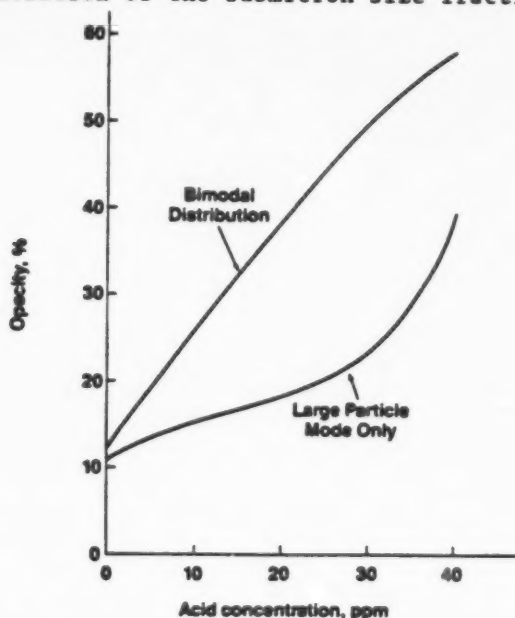


Figure 3. Peak plume opacity with H_2SO_4 concentration.

Because of the dominant nature of the condensable vapor concentration and the concentration and mean size of the submicron particulate fraction, any control approaches for the reduction of the detached plume opacities must be based on them. The plume opacity may be reduced by reducing either or both of these concentrations. Different combinations of the concentration levels for both the vapor phase and fine particulate phase may thus be possible to achieve a desired result.

POSSIBLE CONTROL APPROACHES

As mentioned earlier, both the concentration of the condensable vapor and that of the fine particle fraction of the fly ash present in the stack

gases have a strong influence on the detached plume opacity. Consequently, to achieve a desired peak plume opacity different combinations of these two variables may be selected. A number of simulations were, therefore, carried out to obtain the iso-opacity lines as a function of the acid vapor concentration and the fine particle mode concentration for one set of system parameters. The large particle mode has an additive effect primarily in determining the overall extinction coefficient of the plume. Therefore, these simulation were carried out by setting the large particle concentration equal to zero. All other variable values were those used in the base case simulations described earlier. The results are shown graphically in Figure 4, for small particles at $0.15\ \mu\text{m}$ number mean diameter. Incidentally, a small change in this mean mode diameter can also have some small effect on the plume opacity, as seen in Figure 5, obtained with a mean diameter of $0.1\ \mu\text{m}$. (This figure is shown here only for comparison with Figure 4; further discussion refers to Figure 4.) The opacities indicated in Figure 4 do not include the contribution by the large particles, which may be easily done by considering the in-stack opacity for the total particulate phase. Also, note that other parameters (e.g., water vapor concentration, plume velocity, and ambient conditions) do affect plume opacity, although moderately. The fine particle size distribution is assumed to be lognormal with a geometric standard deviation of 1.3; whereas, the actual distribution may be different. Thus, Figure 4 is not unique: it needs to be reconstructed for any specific case.

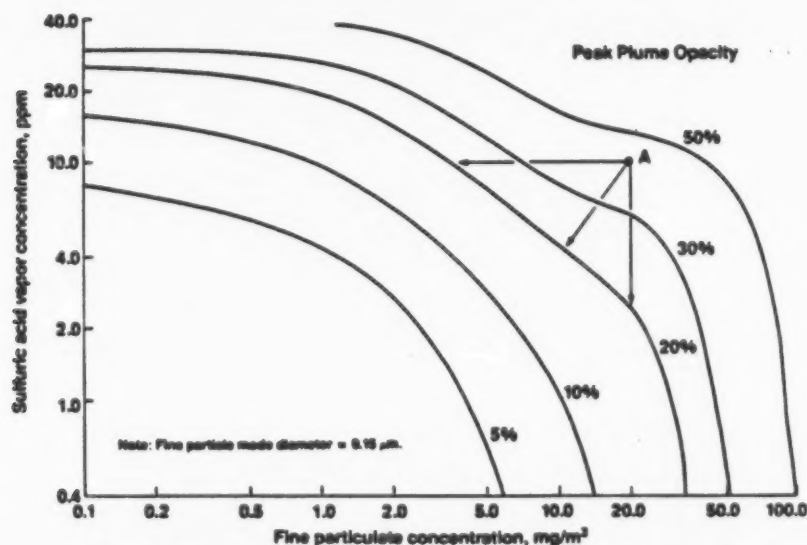


Figure 4. Iso-opacity plot ($0.15\ \mu\text{m}$ diameter).

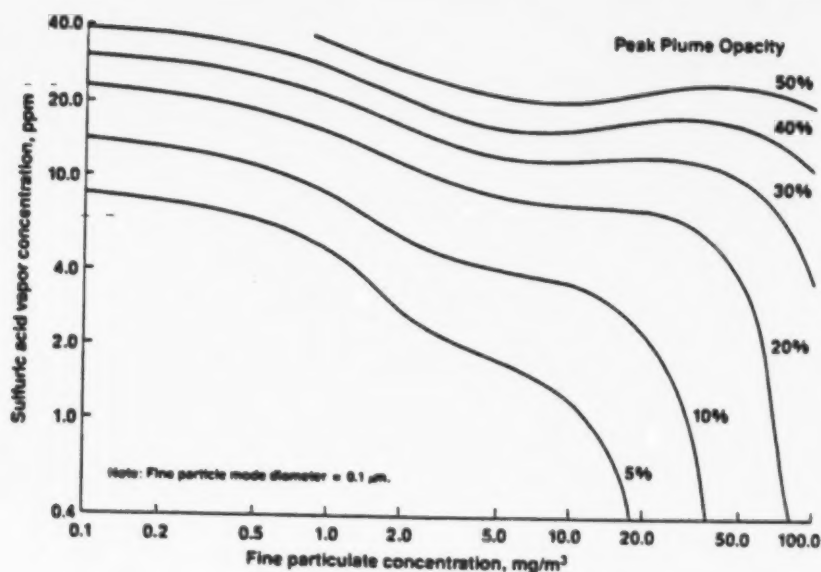


Figure 5. Iso-opacity plot (0.1 μm diameter).

In Figure 4, a given situation in a certain detached plume opacity problem is indicated as a point (e.g., point A in Figure 4 represents 10 ppm of sulfuric acid vapor and 20 mg/m^3 of fine particulate phase in the stack gas). The desired opacity is, however, shown by a line of different vapor and fine particulate concentration combinations (for example, the iso-opacity line of 20% opacity in Figure 4). Several approaches are thus possible to achieve a desired opacity; e.g., by reducing vapor concentration, or by reducing fine particulate concentration, or by both (as shown by arrows from Point A in Figure 4). Reduction in fine particulate concentration does not normally affect the vapor concentration. The condensable vapors may be reduced either in the gas phase (by using a dry reagent to remove vapors or by process modifications) or in the liquid phase (by condensation followed by droplet collection along with other particulate matter). The liquid phase removal of condensable vapors directly affects the particulate concentration as the condensation process generates additional droplets. The gas-phase removal of vapors is also likely to increase, although indirectly, the particulate concentrations. These possible approaches are shown schematically in Figure 6. The iso-opacity lines for a given case identify the concentration reductions required in a given approach. The individual approaches of reducing fine particulate concentration or vapor concentration are considered below for their applicability, advantages, and disadvantages.

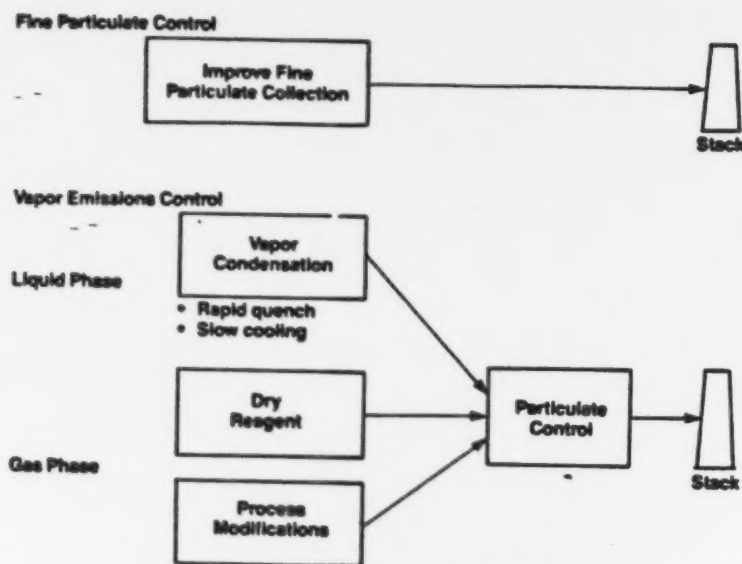


Figure 6. Schematic of possible control approaches.

CONTROL OF FINE PARTICLE CONCENTRATION

The approach of reducing existing fine particulate concentration, if feasible, is probably the simplest. Its feasibility, however, depends not only on the particulate phase concentration and efficiency of the existing particulate collection system, but also on the level of sulfuric acid vapor present. For low vapor concentrations, the opacity may be reduced satisfactorily by a moderate reduction in the fine particulate concentration; but, at a high vapor concentration, even a large reduction in fine particulate concentration may not achieve the desired plume opacity level.

For example, referring to Figure 4, for a 10 ppm acid vapor concentration, a factor of two reduction (from 7 to 3.5 mg/m³) in fine particulate concentration will reduce the opacity from 30 to 20% and may be satisfactory. However, the same reduction in fine particulate concentration for an acid vapor concentration of 20 ppm will reduce the opacity from 50 to 35% and will not be satisfactory. At a 20 ppm acid vapor concentration, further reduction of fine particulate concentration from 3.5 to 1.0 mg/m³ will be required for the desired 20% opacity.

If the existing particulate control facility can achieve the required reduction in fine particulate concentration, either by improving the performance and capacity of an existing unit or by upgrading the facility, such an approach of reducing fine particulate concentration to reduce the plume opacity will be feasible. The main advantages of this approach, if feasible, are the simplicity of the approach, retrofit possibilities to an existing facility, and that the particulate concentration will be reduced without affecting the acid vapor concentration. The applicability of this approach,

of course, depends on the capability of the existing particulate control facility and any possibility of upgrading it if required. By noting the typical fine particulate concentrations at the outlet of high performance particulate control equipment, such as an ESP or a baghouse, tolerable concentrations of acid vapor in the flue gas can be estimated for a certain maximum plume opacity; e.g., 20%.

In a performance evaluation program carried out by Gooch and Merchant (1978), the performance of six ESPs was evaluated. Based on their measurements, average emissions of total particulate matter at the ESP outlets under normal operation were about 16.5 mg/m^3 ; and average emissions of fine particulates were about 3.0 mg/m^3 . In the same study, fine particulate emissions less than 1.0 mg/m^3 were also often noted in some tests. Referring to Figure 4, an average outlet concentration of fine particulates of 3.0 mg/m^3 will allow sulfuric acid vapor emissions of about 8 ppm for a maximum opacity limit of 20%, assuming a 4 to 5% opacity contribution by the large particle mode. If a particular facility can limit the fine particulate emissions below 1.0 mg/m^3 , it can allow as much as 20 ppm of sulfuric acid vapor in the flue gas emissions for a maximum opacity limit of 20%. Note that, based on Figure 4, any further reduction in fine particulate emissions will not allow any significantly higher sulfuric acid vapor emissions in the flue gases. Baghouses typically have lower particulate matter emissions than ESPs. Based on a performance evaluation of a fabric filter by Ensor et al. (1981), total particulate emissions are typically less than 5.0 mg/m^3 . The fine particulate emissions may, therefore, be expected to be less than 1.0 mg/m^3 . Therefore, based on Figure 4, baghouses may typically allow as much as 20 ppm of sulfuric acid vapor in flue gas emissions, for a maximum opacity limit of 20%.

If, in a given situation, the sulfuric acid vapor emissions are such that any possible and reasonable reduction in fine particulate concentration is not adequate to lower the plume opacity to a desired level, sulfuric acid vapor concentration need to be reduced to achieve the desired opacity level. In such a case, reduction of fine particulate concentration alone will not be technically feasible.

REDUCING ACID VAPOR CONCENTRATION

For high sulfuric acid vapor concentrations in the flue gas, it will be necessary to reduce the acid vapor content to meet the opacity requirements. For example, as indicated by the discussion in the previous section referring to Figure 4, if the vapor level exceeds 20 ppm in the flue gas, vapor removal will most likely be required to meet a 20% plume opacity requirement. The vapor concentrations may be reduced in the liquid phase or in the gas phase. In the liquid-phase removal, vapors are first condensed. Based on the rate of gas cooling, the condensation process may be divided in two groups: rapid quenching of the gas phase by direct contact with water, and slow cooling of the gas as in a heat exchanger/condenser. The vapors may be removed in the gas phase by using a reagent that selectively reacts with the vapor species. It may also be possible to reduce the gas-phase vapor concentrations by modifications of processes responsible for vapor generation. The possible approaches for reduction in the vapor concentration are:

- 1) Rapid quenching of the flue gas in a water/flue gas contactor--a packed bed, for example.
- 2) Slow cooling of the flue gas in a heat exchanger.
- 3) Injection of a dry active reagent such as nacholite.
- 4) Process modifications to reduce generation of acid vapors.

One principal difference in approaches for vapor removal, as compared to those for particulate removal, is that all above-mentioned vapor removal processes are likely to produce additional particulate loadings in the flue gas; whereas, reducing the fine particulate concentration is not likely to affect the acid vapor concentration. Approaches (1) and (2) condense the vapors, directly converting them into particulate matter. In approach (3), fine reagent powder is injected into the flue gas to form the additional particulate loading. Approach (4), if based on using less excess air/oxygen in the combustion process, may produce incomplete combustion, thereby increasing the particulate loading. Therefore, any consideration of these approaches must also include adequate removal of the additional generated particulate matter. Since the submicron particles are the hardest to collect in a particulate control device, attempts must be made to minimize the fine particulate concentration during vapor removal, especially by condensation.

An experimental program was recently undertaken by Yung et al. (1984) to study the condensation of sulfuric acid vapors. In this program, vapor condensation was studied under different operating conditions using three process configurations: rapid quenching, dilution cooling, and slow cooling. In all experiments, the size distribution and concentration of the aerosols generated by the condensation process and the amount of the acid vapor condensed were measured as a function of different operating parameters. These results, especially for the rapid quenching and slow cooling experiments, indicate the operating conditions which minimize the fine particulate formation in these condensation processes.

These four approaches are considered separately for their feasibility and applicability.

1. Rapid quenching of the flue gas. This process inherently generates a large number of sulfuric acid vapor droplets by homogeneous nucleation, and attention must be given to minimize submicron particulate formation. In this approach, the flue gas may be directly contacted by quench water in a packed bed or spray column. In the experiments by Yung et al. (1984) simulated flue gas containing known amounts of sulfuric acid vapor, water vapor, fly ash, and SO_2 was quenched in a randomly packed bed of berl saddles by a known countercurrent flow of quenching water. These experiments indicate that higher acid vapor concentration, lower quench water temperature, and higher water vapor concentration reduce the proportion of fines in the total particulates. Their experiments also show that almost all of the acid vapors can be condensed in this process, and that about 50% of the total amount of acid is condensed in the bed. The fly ash, although somewhat effective in scavenging the acid droplets, did not prevent the acid vapor from nucleating into fine droplets.

Condensation of 1 ppm of acid at 120°C generates about 3.5 to 4 mg/m³ of particulates. Thus, based on the amount of sulfuric acid vapors in the flue gas and the measured size distribution of the generated aerosols in the above mentioned experiments, the amount of total and fine particulates generated in this process may be estimated. This information may then be used to determine the additional particulate loadings and the adequacy of the existing particulate control equipment to handle them. Since this process removes almost all of the acid vapors, the in-stack particulate loading is measureable by standard stack sampling techniques (e.g., Method 5), and the correlation between the in-stack particulate mass and the opacity is restored.

This process is suitable for high acid vapor concentrations as indicated by experiments. This is not actually a limitation, since such vapor removal will most likely be undertaken for high acid vapor concentrations anyway. This process also favors high flue gas moisture content which allows growth of the condensed droplets in the scrubber. It may be worthwhile even to consider injecting waste steam, if available, into the flue gas to increase the water vapor concentration. This essentially is the Flux/Force Condensation System suggested by Calvert et al. (1975) to remove fine particulates. Since this process removes almost all of the acid, based on Figure 4, up to 30 mg/m³ of fine particulate may be allowed in the emissions without violating the 20% opacity limit. This fact may allow an existing particulate control facility to handle the additional particulate loading. Additional factors that must be considered in evaluating the feasibility of this process are: additional space and ducting required for the scrubber equipment, the material of construction for the scrubber to handle the condensed dilute acid, the possibility of corrosion in subsequent particulate control equipment due to collection of acid droplets, the additional pressure drop across the scrubber, and the need for reheating the flue gas before stack exit.

The ability of the existing particulate control equipment to handle the greater particulate loading at lower temperature, especially additional acid droplet loading, is very important as it can have direct impact on the economics. If the existing equipment is not suited for acid droplets and/or lower temperature, it may be necessary to install the quench scrubber following the existing primary particulate control system. Additional smaller size particulate control equipment capable of handling acid droplets will then be needed to remove generated acid droplets.

2. Slow Cooling. This process relies on controlled cooling so that the gas temperature is lowered below the dew point (to commence condensation), but held above the nucleation temperature (to prevent any homogeneous nucleation from generating fine droplets). This method is better suited for lower acid vapor and water vapor concentrations, as this allows a better control on the cooling and condensation process. The presence of fly ash is very important in this process, as it allows a large surface area on which the vapors can condense. In the experiments by Yung et al. (1984), higher acid vapor concentrations resulted in large numbers of fine particulates by nucleation. Thus, for high acid vapor concentrations, applicability of this process over rapid quenching is questionable. Also, this will be a very difficult system to design because close control of gas cooling and temper-

ature is required. One advantage of this process is that, if successful, large amounts of acid will condense on existing fly ash particulates and condenser walls, thus generating only a small amount of additional fine and total particulate loading.

For low acid vapor concentration, this process may be preferable to rapid quenching; however, in such a case, plume opacity may be controlled adequately by reduction in fine particulate concentration alone. For low vapor concentration, reducing fine particulate matter is better suited than slow cooling due to process simplicity and the possibility of retrofit application. As for rapid quenching, attention must also be given to other important factors; e.g., additional space and equipment, the material of construction, corrosion in the particulate control system, the additional pressure drop, and the need for reheat. The ability of the existing particulate control system to handle acid droplets must also be considered.

3. Injection of Dry Reagent. One advantage of this approach is that it does not generate additional fine particulate matter. The dry powder used is usually coarser, with a mean diameter of 20 to 100 μm . Thus, although this process will generate additional particulate loading, it will not be difficult to collect. Nacholite has been successfully used to remove SO_2 from flue gas with moderate efficiency, and the sulfuric acid may be expected to be even more reactive than SO_2 . Such dry injection processes for SO_2 have typically been used with a baghouse, and the cake deposit on the fabric bags containing the reagent is thought to provide the necessary gas-phase/reagent contact. Thus, such a process may be better suited in conjunction with existing baghouse facilities. Spray drying has also been used to remove SO_2 ; however, such a process leads to quenching of the gas phase which in this case will produce many fine droplets which will not be scavenged by the reagent slurry droplets.

In evaluating this process, the ability of the reagent (e.g., nacholite) to react with a relatively low concentration of acid vapor must be determined. The concentration of SO_2 to which this process has been applied is usually of the order of 500 ppm. Whether such reactivity will be observed with 50 ppm or less of sulfuric acid vapor must be determined experimentally. No such data are known to be available. If the dry reagent is found to be reactive enough to remove small quantities of sulfuric acid vapor efficiently, this process will be a good alternative to the above mentioned condensation processes.

4. Process Modifications. This control approach is aimed at reducing the amount of sulfuric acid vapor generated in the fuel combustion process, rather than removing the generated acid vapors. Several studies have evaluated the effect of different operating conditions on the sulfuric acid vapor emissions: they indicate that the percent of sulfur in fuel and the percent of excess air/oxygen used in the combustion process predominantly determine the acid vapor emissions. Additionally, oil fired units are generally found to produce greater emissions of the acid vapor as compared to coal fired units. This difference is generally attributed to the presence of vanadium in the fuel oil which serves as a catalyst in the SO_2 conversion reaction.

Fuel combustion produces almost quantitative amounts of SO_2 . The conversion of SO_2 to SO_3 and primary sulfates is predominantly a function of excess oxygen levels in the combustion gases and the presence of catalysts such as vanadium. Thus, for a given combustion facility almost quantitative reduction in SO_2 and SO_3 emissions can be achieved by reducing the sulfur content of the fuel. The use of low sulfur fuel, as a control option, is limited, however, by the cost and availability of such low sulfur fuel. Note that this control option is mandatory in some parts of the United States; e.g., New York City.

Emissions of SO_3 from utility and industrial boilers typically vary from 1 to 6% of total SO_x depending on percent excess air/oxygen. The available data in the literature, however, show lots of scatter (Homolya and Cheney, 1978; Hunter and Engel, 1978). Theoretical computation by Hunter and Engel (1978) indicated about 3% conversion at 2% excess oxygen with a leveling in the percent conversion at higher oxygen levels. At levels of oxygen less than 2%, almost linear variation of flue gas SO_3 concentration has been observed with respect to excess oxygen (e.g., Dietz et al., 1978). Thus, based on theoretical calculations and field measurements, it appears to be possible to reduce SO_3 emissions by reducing excess air used in the combustion process. Successful operation of a combustor with low excess air requires that the oxygen in the flue gas be maintained at levels below 0.2%. Attention must be given in such operation to possible additional particulate phase formation due to incomplete combustion. Such operation requires precise control of the fuel/air ratio in all parts of the combustion system. Such a low excess air operation is more practical for oil-fired systems; the required technology is not available for coal-fired systems (Levy et al., 1978).

Levy et al. (1978) also investigated effects of other combustion modifications (e.g., staged combustion and flue gas recirculation) on SO_3 formation. In staged combustion, even an enhancement of SO_3 formation was found to be possible; whereas, other combustion modifications were found to have little or no effect on the SO_3 concentration in the flue gas. Thus, besides using a low sulfur fuel, operation at low excess air appears to be the only process modification capable of reducing the amount of SO_3 generated. This approach is also probably suitable only for oil-fired units.

CONCLUSIONS AND RECOMMENDATIONS

The problem of "detached plume" opacity caused by condensation of sulfuric acid vapors was analyzed using basic principles involved in this physical process. The concentration of condensable vapors and the concentration of fine particulate matter in the flue gas are the most important parameters influencing the maximum downwind plume opacity. Control strategies to reduce the detached plume opacity must, therefore, be based on reducing the fine particulate concentration in the flue gas, or the condensable vapor concentration, or both. Using the "PLUME" model, iso-opacity lines may be constructed as a function of the concentrations of the condensable vapors and those of the fine particulate phase, for a given situation taking into account all other system, operating, and ambient meteorological

parameters. The iso-opacity lines provide a graphical indication of possible control approaches.

In general, for low sulfuric acid vapor concentrations, typically below 20 ppm, reduction of fine particulate concentration alone may provide adequate control of the plume opacity. For higher sulfuric acid vapor concentrations, reduction in acid vapor concentration will generally be necessary along with adequate particulate control to reduce the plume opacity. Three specific approaches may be possible in reducing the acid vapor concentration: use of low sulfur fuel, rapid quenching of the flue gas in a packed bed scrubber, and combustion under low excess air conditions. In the latter two approaches, especially in flue gas quenching, attention must be given to additional particulate-phase formation which must be handled by the existing particulate control systems. The low excess air operation may be suitable only for oil-fired systems. Injection of dry reagent powder may also be a feasible approach to remove acid vapors in the gas phase, although experimental data are needed to verify it.

In summary, this paper outlines the boundaries of the condensation aerosols problem. Established control technology to reduce the detached plume opacity problem, especially to reduce vapor concentrations in the flue gas, does not exist at this time and must be developed for effective control of plume opacity. Possible approaches in that direction are outlined in this paper.

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RELATIONSHIP OF OPACITY TO THE LOSS MECHANISMS IN ESPS

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ABSTRACT

Electrostatic precipitators have been subjected to extensive testing for a variety of reasons including testing for compliance, determining relationships of mass concentration to plume opacity, troubleshooting to improve performance, and gathering information to develop design models. The limited objectives of many of the studies may limit the interpretation and usefulness of the reported results.

This paper describes the use of electrostatic precipitator models and key measurements, such as opacity, to obtain an objective assessment of plant operation. The models show clear trends in the importance of sneakeage and rapping reentrainment on the emissions and opacity.

This paper has been reviewed in accordance with the U.S. Environmental Protection Agency's peer and administrative review policies and approved for presentation and publication.

INTRODUCTION

The interpretation of field data from precipitators has always been a critical part of the engineering of emission control equipment. Traditionally, information such as power, voltage-current characteristics, and inlet/outlet mass concentration has formed the bulk of technical information. However, in recent years, computer simulation models have facilitated understanding of the precipitator from a mechanistic standpoint. Computer models were used in the development of this paper to show the value of combining independent measurements such as opacity and mass emissions to aid in precipitator troubleshooting.

An additional emphasis for studying the effects of electrostatic precipitators on opacity is the importance of mass-opacity relationships in air pollution regulation. Stationary sources are regulated by both mass emissions and plume opacity limitations. Situations have existed where the source has been in compliance with the mass emission regulation but the emissions have exceeded the plume opacity limitation for the particular industry. Recently, the legal mechanisms have been available to develop special opacity limitations for specific sources. As a result, several states have source testing programs to relate opacity to mass emissions. In addition, the in-stack transmissometer is the only continuous monitor for particulate matter, and interest has been shown in calibrating the instrument as a mass emission monitor (for example, see Reference 1).

COMPUTER MODELING

ELECTROSTATIC PRECIPITATORS

In recent years, good computer simulations of electrostatic precipitators have been developed. Examples include the Southern Research Institute/Environmental Protection Agency (SoRI/EPA) model described by Mosley et al.²

and the Research Triangle Institute (RTI)/EPA model described by Lawless et al.³ These EPA-sponsored models are based on fundamentals of particle charging and migration to the collector plates.

The development of these computer programs has received considerable attention and therefore will not be covered in detail here. An important limitation is that not all of the significant parameters such as rapping entrainment and sneaking can be predicted directly. Thus, empirical data are used to correct the model. However, experience has shown that rapping entrainment is usually 10 percent per section. Also, the material rapped from precipitator plates in puffs seems consistently to have a mean particle diameter of about 6 micrometers and a standard deviation of 2.5.

OPACITY

The stack transmissometer or opacity meter is a common instrument for real-time measurement of aerosol emissions. Opacity meters are often used to perform adjustments in the rappers to optimize the operation of the precipitator.⁴

The electrostatic precipitator models have program modules that allow the computation of the expected in-stack opacity. Both static and dynamic models have been developed with this capacity. These models have allowed investigation of the factors that affect opacity. Two objectives have motivated the opacity study: to strengthen the use of opacity as a troubleshooting tool and to explain stack testing data used to develop mass-to-opacity correlations.

The relationship of plume opacity (O_p) to mass concentration is given by the following equations where opacity is 1 minus light transmittance.

$$O_p = 100(1 - I/I_o), \quad (1)$$

where I/I_o is the light transmittance over the stack diameter L (m) and is related by the Bouguer law to the plume properties.

$$I/I_o = e^{-b_{ext}L}, \quad (2)$$

where b_{ext} (m^{-1}) is the extinction coefficient. The extinction coefficient is related to mass concentration by taking the following ratio:

$$S_m = b_{ext}/W, \quad (3)$$

where S_m (m^2/g) is the mass extinction coefficient, and W (g/m^3) is the particulate mass concentration under actual conditions. Thus:

$$O_p = 100(1 - e^{-S_m W L}). \quad (4)$$

The mass extinction coefficient for spherical particles is given by:

$$S_m = \frac{\int_{d_1}^{d_2} \pi Q D^2/4 N(D) dD}{\rho \int_{d_1}^{d_2} \pi/6 D^3 N(D) dD} \quad (5)$$

where Q is the light extinction efficiency, D is the particle diameter (μm), $N(D)$ is the particle size distribution by number ($\#/m^3$ between D and $D + \Delta D$), and ρ is the particle density (g/cm^3).

The Q is computed with Mie theory and depends on the particle diameter, the refractive index, and the wavelength of light.⁵ Thus, the mass extinction coefficient is sensitive to the particle size distribution. The particle size distributions computed by the electrostatic precipitator are then used in Equation 5 to compute S_m and the resulting opacity. The larger the S_m , the greater the opacity expected for a given mass concentration.

ELECTROSTATIC PRECIPITATOR-OPACITY STUDY

The ESP-opacity model can be used to predict how opacity will be affected by certain precipitator parameters such as specific collector area and rapping. The predicted behavior (in terms of opacity and S_m) can then be compared to field data. The present work is an extension of work presented previously.⁵

Conditions typical of an electrostatic precipitator controlling high sulfur coal with an ash of optimum resistivity were used in the model. These conditions seem to be particularly well predicted by the model used.³ The case study is shown in Table 1. The changes in opacity with respect to collection area and rapping were of particular interest. The collection area or the opacity-mass relationship is important because a popular method of obtaining variability in the outlet mass concentration opacity is to

TABLE 1. SUMMARY OF ESP PARAMETERS

Parameter	Value
Wire-to-plate spacing	0.127 m
Wire radius	2.54×10^{-10} m
Area per section	1,477 m ²
Sections	4
Gas flow rate	117 m ³ /s
Total specific collection area	50.5 m ² /m ³ /s (256 ft ² /ft ³ /min)
Temperature	433 K
Pressure	1.0 atm
Length of section	9.76 m
Resistivity	2×10^8 ohm-cm
Applied voltage (depending on section and operating conditions)	Up to 55 kV
Current density (depending on section and operating conditions)	Up to 4×10^{-4} A/m ²
Sneakage/section	3%
Rapping reentrainment/section (An average cold-side puff size distribution from Gooch and Marchant ⁶ was used.)	0%, 10%, 20% $D_g = 6.0 \mu\text{m}$, $\sigma_g = 2.5$
Sigma (standard deviation of gas velocity)	0.25
Inlet size distribution	
Mass mean diameter	20 μm
Geometric standard deviation	4
Mass concentration	9 g/m ³
Particle density	2.27 g/cm ³
Refractive index	1.525 0.05 i

conduct tests with the field shut off. This experimental procedure is justified on the basis that the most common failure mode and its effect on opacity is simulated. A second very important parameter is the amount of rapping emissions. A previous paper⁵ demonstrated that rapping may overwhelm the effects of inlet particle size distribution.

The modeling results are summarized in Table 2 and shown in Figures 1 and 2. In Figure 1, the mass extinction coefficient is plotted as a function of collector area for conditions of rapping reentrainment. The fly ash from the boiler typically has a reasonably narrow range of S_m .⁷ However, as the particles move through the precipitator, the size distribution is changed. Two distinct mechanisms are at work in modifying the particle size distribution: the large and small segments of the distribution are being removed more efficiently than the 0.5-to-1- μm region (the collection window) causing the particle size distribution to be reduced in size, and material is being reentrained during rapping at about 6- μm diameter. The relative contribution of these two mechanisms will shape the size distribution emitted from each section. The size distribution will in turn affect the opacity/mass concentration relationship.

The most likely curve for rapping reentrainment is the 10-percent line. Dry wall electrostatic precipitators would not have zero reentrainment. The 20-percent reentrainment is very heavy and represents a poorly adjusted machine, although reentrainments as high as 80 percent have been reported.⁷ The S_m for the portion of the fly ash distribution in the particle size region of minimum collection efficiency (particle diameter = 1 μm) is 2.4 m^2/g , while the S_m for the rapping reentrained material (geometric mean diameter = 6 μm , geometric standard deviation = 2.5) is about 0.4 m^2/g . For the 20- μm fly ash, both mechanisms will increase the magnitude of S_m . However, as demonstrated in a previous paper,⁵ if the incoming ash is less than 6 μm in diameter, the S_m may be reduced in magnitude depending on the prominence of rapping reentrainment.

In Figure 2, the numerical results have been replotted with S_m as a function of penetration through the electrostatic precipitator. Keep in mind that, for the simulation with a constant inlet concentration, penetration will be proportional to mass concentration. These results demonstrate that the greater the efficiency (lower outlet concentration) the larger the

TABLE 2. RESULTS OF ELECTROSTATIC PRECIPITATOR SIMULATION

Section	Inlet	1	2	3	4 (outlet)
Cumulative area, m ²	0	1,477	2,954	4,431	5,908
Cumulative SCA, m ² /m ³ /sec	0	12.6	25.2	37.9	50.5
<u>No rapping reentrainment</u>					
Mass concentration, g/m ³	9.16	0.279	0.0595	0.0156	0.0046
Opacity (over 6 m), percent	99.9	73.19	44.84	14.8	5.88
S _m , m ² /g	0.285	1.78	2.84	3.89	4.79
Penetration (fractional)	--	0.0305	0.0065	0.0017	0.0005
<u>10% Rapping reentrainment</u>					
Mass concentration, g/m ³	9.16	0.761	0.129	0.0266	0.00701
Opacity (over 6 m), percent	99.9	84.7	50.6	15.9	6.13
S _m , m ² /g	0.285	0.935	1.54	2.44	3.41
Penetration (fractional)	--	0.083	0.0141	0.0029	0.0008
<u>20% Rapping reentrainment</u>					
Mass concentration, g/m ³	9.16	1.24	0.312	0.0824	0.0232
Opacity (over 6 m), percent	99.9	91.3	53.6	22.6	8.8
S _m , m ² /g	0.285	0.744	0.93	1.18	1.51
Penetration (fractional)	--	0.1356	0.0341	0.009	0.0025

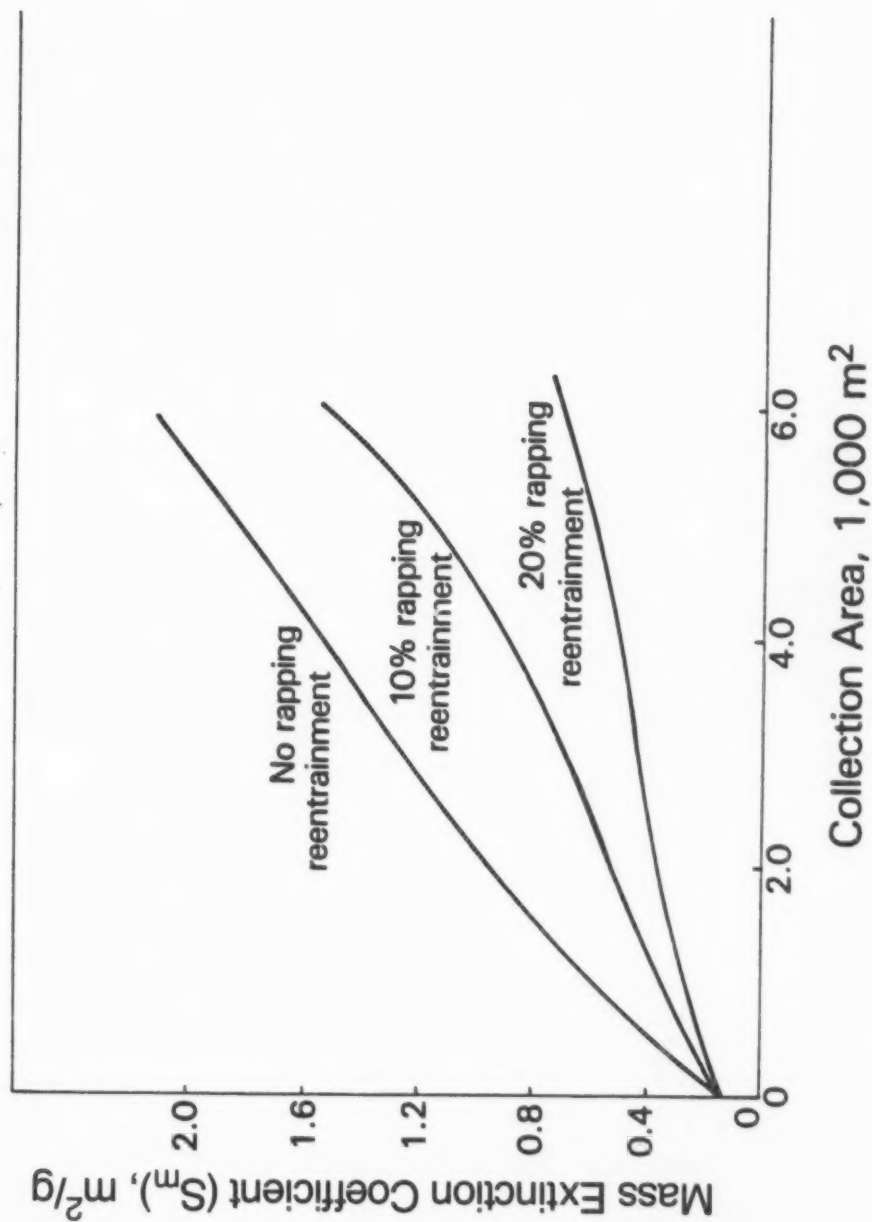


Figure 1. Mass extinction coefficient as a function of precipitator collection area.

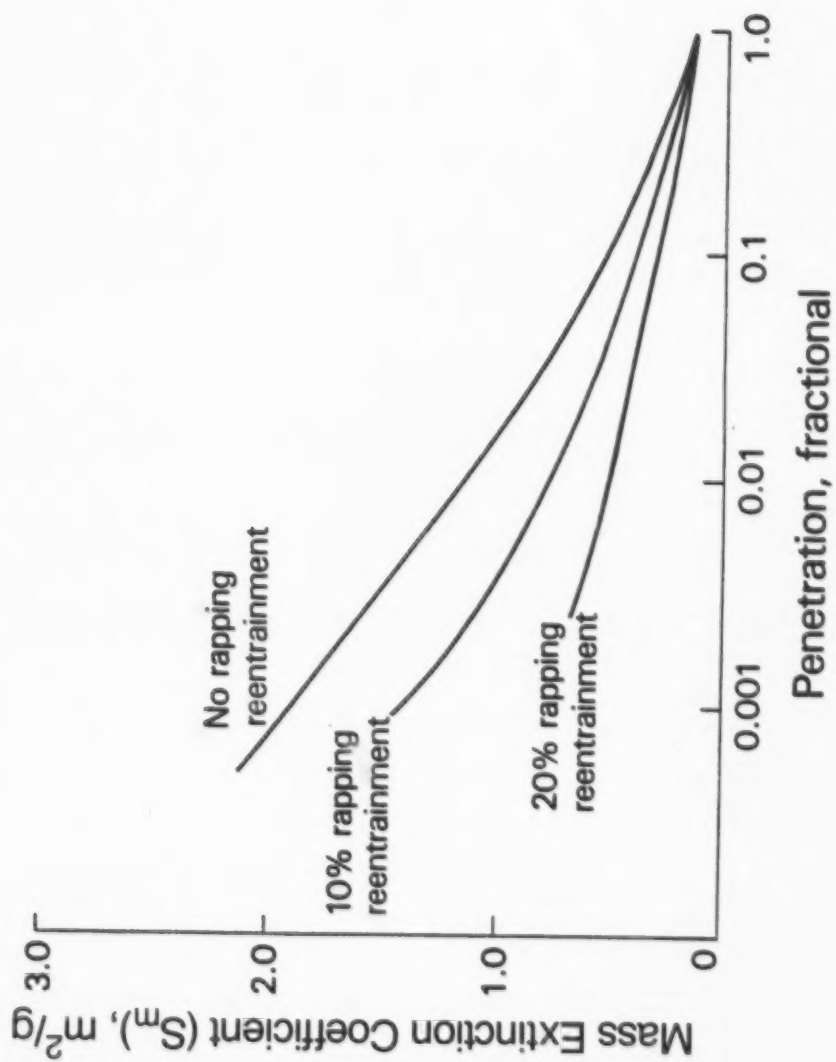


Figure 2. Mass extinction coefficient as a function of penetration.

measured S_m . An exception is for cases of high rapping entrainment losses as indicated by the 20-percent rapping reentrainment curve. Two basic conclusions can be drawn from this curve: first, the assumption often made that the size distribution must be constant (hence a constant S_m) to obtain a correlation between opacity and mass probability is shown not to be the case in this simulation. In particular, when fields of the electrostatic precipitator are shut off to obtain a wide range of concentration, S_m will vary as indicated. The approach often used to reduce mass concentration-opacity data is to plot the logarithm of transmittance ($1-Op/100$) as a function of mass concentration following Equations 2 and 3.¹ If S_m changes with concentration, a curved rather than straight line will result. The cases from the simulation are plotted in logarithmic linear coordinates in Figure 3. The case with no rapping reentrainment has the greatest curvature because of the significant influence of the precipitator removal mechanisms on modifying the particle size distribution. The most likely case of 10 percent rapping reentrainment with the boldface line has some curvature. Quite likely in field data, the curvature would be not noticed because it would be masked by experimental error. A more realistic criterion for correlation of opacity with mass concentration is that the electrostatic precipitator modify the particle size distribution in a consistent manner as a function of mass concentration. Second, computation of the mass extinction coefficient, S_m , is a useful way of determining the operating mode of the precipitator. If the measured S_m is less than $0.5 \text{ m}^2/\text{g}$, quite likely the precipitator has very heavy reentrainment and should be repaired. Thus, by using opacity mass measurements and knowledge of the operation of the precipitator, insight can be gained into the operation of the precipitator with limited test work.

FIELD DATA

The next step in testing the hypothesis developed with the ESP-opacity model is to evaluate field data. Unfortunately, when the objective is to obtain data to characterize the electrostatic precipitator, obtaining accurate opacity data is usually neglected and vice versa. For the purposes of this paper, only sample data will be considered to demonstrate qualita-

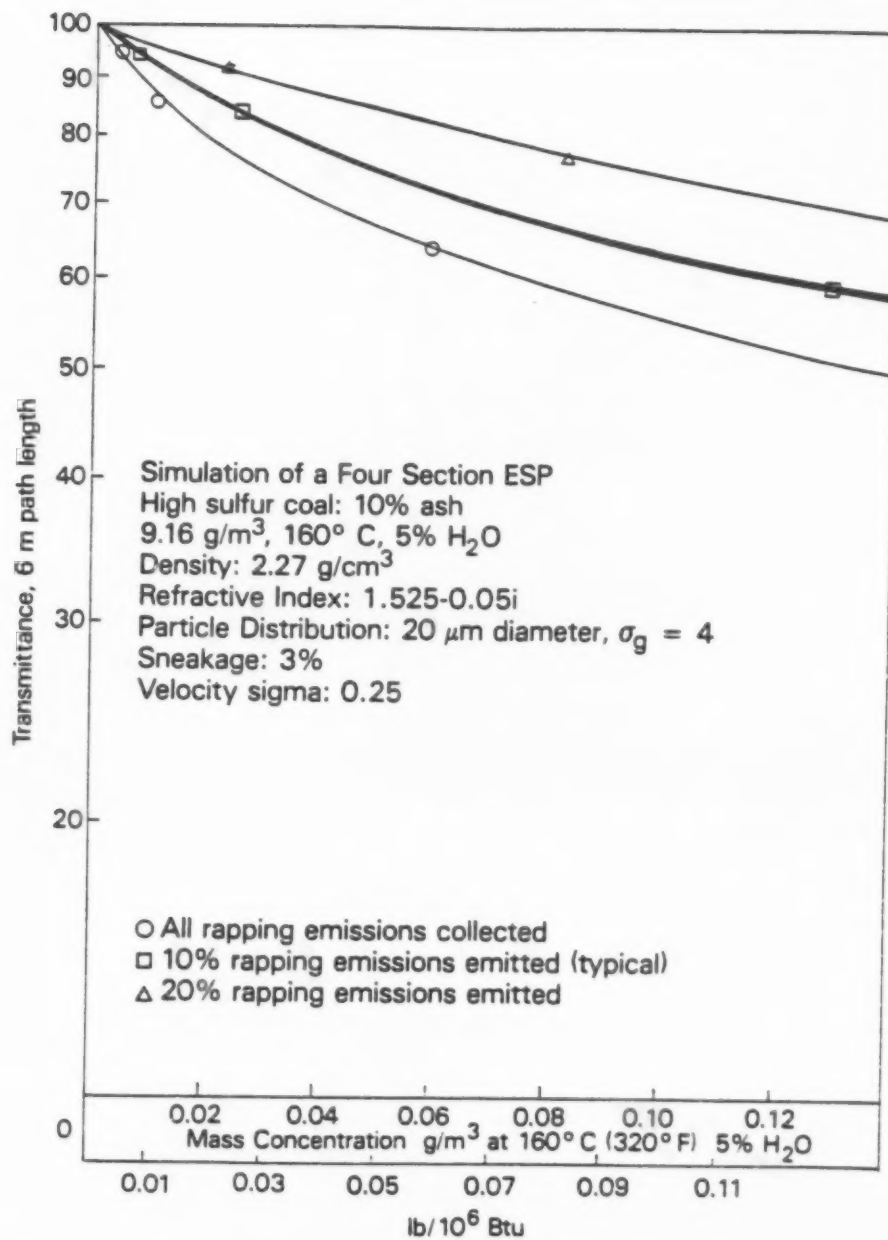


Figure 3. Light transmittance as a function of mass concentration for the computer simulation case.

tively the potential of the approach. A typical example of field-determined mass extinction coefficients is shown in Figure 4 from Reference 8. A high correlation of mass and opacity would be expected because of the systematic variation of S_m with mass concentration and, in fact, a correlation coefficient of 0.98 was reported. If rapping reentrainment or other nonidealities were important from time to time, a poor correlation of mass and opacity would be expected. The shape of the mass extinction coefficient S_m as a function of outlet concentration suggests that the performance of the precipitator is following the moderate rapping reentrainment curve. If the curve were invariant with concentration, heavy rapping would be suspected.

Obviously these sample data are highly suggestive, but are not conclusive evidence of the validity of the approach. As a followup to field test and prove the concept, two approaches are suggested: a careful search of the rather limited published data, and tests designed to obtain complete data sets that can be used to test the concept.

CONCLUSIONS

The use of opacity data to interpret the operation of electrostatic precipitators is a neglected area of field data analysis. The mass extinction coefficient can be used to deduce the mode of operation in the precipitator.

If the precipitator has excessive rapping emissions indicated by a small, constant S_m , it would be worthwhile to investigate the rapping operation. If the S_m changes with outlet concentration, then rapping reentrainment is likely not a problem. If the precipitator is primarily limited by low collection efficiency indicated by an S_m that changes with outlet concentration, then an effort should be made to improve the collection efficiency. From a research standpoint, the analysis provides useful insight into limitations of existing electrostatic precipitator designs.

Field Tests—Georgia Power
 Branch 1 250 MW
 Branch 2 319 MW
 Branch 3 481 MW
 Branch 4 490 MW
 51 tests

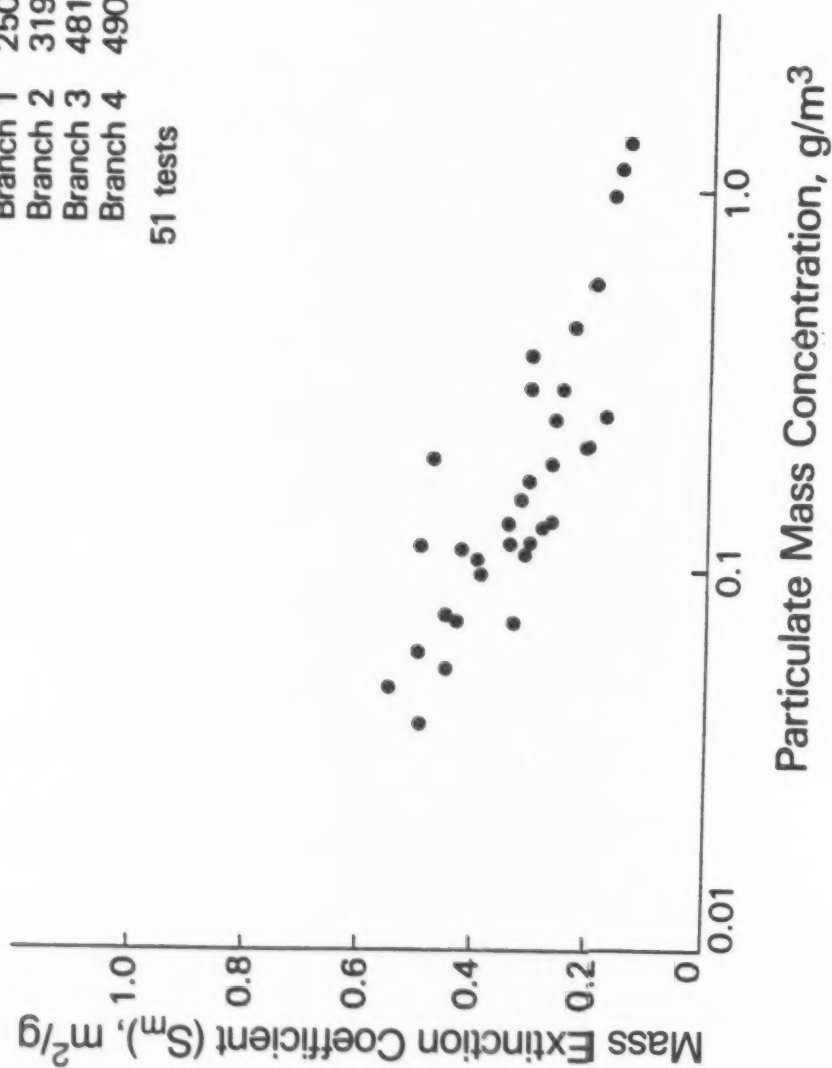


Figure 4. Mass extinction coefficient of field data.⁸

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STACK OPACITY MODEL

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ABSTRACT

Phase III of a particulate emissions and stack opacity characterization program for oil-fired boilers is currently in progress. Phase I had to do with characterizing particulate emissions while phase II had to do with the effects of various fuel and backend additives. Phase III has to do with the generation of sulfate species and prediction of stack opacity. A model to predict opacity from the boiler design, operation and oil properties was developed.

As part of the investigation, an intensive test series was conducted on a face-fired utility boiler which included simultaneous sampling at the air heater inlet and outlet locations. Measurements of particulate mass, particle sizing, and sulfur trioxide (SO_3) were taken during steady state low load, transient load (period in which full load is just reached), and steady state full load conditions. Correlations of opacity with generation of SO_3 were derived from the data. During transient conditions, the increases of SO_3 emissions were shown to emanate from the boiler surfaces rather than from the air heater surfaces as previously thought by the investigators. The higher generation of SO_3 during load increases, responsible for the so-called "morning opacity" is believed to be due to vanadium compound transformations associated with convective heat transfer surface temperature increases. This paper presents some aspects of the modeling concepts and some results of the testing.

INTRODUCTION

Since 1978, the Consolidated Edison Company of New York (Con Edison) has been engaged in an applied research program to evaluate the properties of particulate emissions and stack opacities from oil-fired boilers. Phase I investigated the characteristics of particulate emissions in terms of mass emissions, particle size distributions, composition and morphology (1) while phase II studied the effects of various types of fuel additives and backend conditioners (2). Phase III, initiated in 1982, focuses on the effects of particulate and SO_3 emissions on the stack plume opacity.

When moderate to high sulfur content oils are burned there is usually an associated stack plume due to a combination of particulate species and condensed SO_3 . The particulate species are formed from metals in the fuel ash, sulfate species from the sulfur in the oil and carbonaceous species formed by incomplete combustion. The SO_3 is generated from the sulfur in the oil. A small portion of the SO_2 (1 to 5 percent) is further oxidized to SO_3 . This occurs in the furnace section and downstream in the convective heat transfer sections of the boiler by catalytic oxidation. Of particular interest is the increased opacity which occurs when load is increased on cycling units. After full load is reached, the increase in opacity takes several hours to dissipate to a steady opacity level. It is believed, on the basis of this study, that this transient is due to additional SO_3 generated by vanadium specie decomposition.

The project is considered important since it is expected that residual oil usage as a utility fuel will continue throughout the 1980's and 1990's especially in plants that are situated in or near populated cities. Further, the residual oil properties over the last several years have been deteriorating. This is primarily due to improved refining processes which favor production of distillate oils thus creating lower quality residuals. Therefore, it is expected that greater emissions and stack opacities will become evident even for lower sulfur oils.

STUDY OBJECTIVES

The Phase III study objective is to develop a computer model which can predict stack plume opacity from boiler design characteristics, fuel properties and operational variables. Ideally, no emissions testing will have to be performed. The model calculates the associated particulate and SO_3 emissions and uses light scattering theory to predict the stack opacity. A second major objective is to calibrate and verify the model by conducting emissions testing on a utility boiler. The emissions testing were conducted on a nominal 350 MW face fired boiler in the Con Edison system.

PROJECT PARTICIPANTS

The sponser for this phase of study was Con Edison. The overall model development and the verification test program were conducted by KVB Incorporated. Sub-models in the overall model were developed by various other companies. The particulate and sizing model was developed by KVB, the SO_3

generation model by Radian Corporation, and the opacity model makes use of the Research Triangle Institute's (RTI) light scattering model while emission test data were reduced by RTI through their opacity model.

MODEL DESIGN

The computer model consists of 4 major subroutines: a boiler heat transfer surface and thermal analysis subroutine, a particulate emission and size distribution subroutine, an SO_3 formation subroutine and a condensation and light scattering opacity subroutine. Figure 1 shows the basic design of the model. Input to the model consists of boiler design information, operating variables and fuel properties. The boiler design information currently is in the form of a subroutine which includes the geometry of the heat transfer surfaces and a set of curves which relate the surface temperatures of these sections to unit load. These data are required for calculating the SO_3 emissions at various load levels.

The operating variables include load level and boiler O_2 . The load can be input as either steam flowrate or turbine/generator output but must be consistent throughout the model. Fuel properties include sulfur content, heating value, ash and vanadium contents. The model calculates the particulate emissions and particle size distribution. The particulate emissions are based on empirical data taken from various sulfur content oil operation in different type utility boilers. The calculation is based on the size of the boiler, sulfur and ash content of the oil, O_2 level, operating load and furnace type. The particle distribution is based on generic data as obtained on different utility boilers over a range of operation (1).

The SO_3 emissions are calculated by three different mechanisms; furnace homogeneous conversion to SO_3 , catalytic conversion over the heat transfer surfaces and vanadium compound decomposition during transient operation. The catalytic and transient SO_3 formations are arrived at by using a plug flow model throughout the convective passes of the boiler. Gas properties are calculated at each pass to determine the associated Reynolds number and residence time.

The model currently does not include the effects of particulate cleaning nor SO_2 scrubbing equipment on the final stack emissions. These can be included by simply adding subroutines prior to calculating the final properties at the stack.

The gas properties and the emission levels are corrected to stack conditions. These corrections include stack temperature and stack diameter. The particulate emissions which are originally generated as mass emission distributions are converted to particle number distributions. The opacity is then calculated separately for the particulate emissions and the SO_3 emissions based on light scattering theory. The opacity is arrived at by a complicated summing routine of the particle size distributions. The computer program solves 50 integrations per log decade of particle diameter ranging from 0.01 to 100 microns. Currently, the model assumes a peak particle diameter and a size distribution for the SO_3 emissions in calculating the

opacity. It does not include particle growth, agglomeration and meteorological effects. The final version will include the RTI/EPA opacity model (3). Comparisons of the test data from this program, however, were made through the RTI/EPA model.

SO₃ FORMATION MODEL

This model determines SO₃ formation rates by three different mechanisms: homogeneous reaction in the furnace zone, heterogeneous reaction across the convective heat transfer surfaces and decomposition of vanadyl sulfate (VOSO₄).

Homogeneous Reaction

Although thermodynamic equilibrium favors complete conversion of SO₂ to SO₃ as the gases pass through the entire boiler system, the high temperature and short residence times cause the reaction to be "frozen" at low conversions. In this model, a simple approach is used to describe the oxidation process. It is assumed that the reaction proceeds according to equilibrium at temperatures observed in the furnace zone and no further reaction takes place thereafter. Since the gas temperatures are high in this region, reaction rate will be high and equilibrium concentrations of SO₃, SO₂ and O₂ are likely to exist at these conditions. Compared to the furnace zone, lower temperatures and shorter residence times justify this assumption. The homogeneous SO₃ formation increases with increasing O₂ and SO₂ concentrations and with decreasing furnace temperature.

Catalytic SO₃ Formation

Vanadium (V) in oil-ash deposits is a good catalyst for the conversion of SO₂ to SO₃ (4). Of the vanadium compounds normally found on tube deposits, V₂O₅ is the only one which is catalytic to SO₂ oxidation. Other predominant vanadium compounds found in fly ash are vanadyl sulfate (VOSO₄), and vanadium tetroxide (V₂O₄). Catalytic oxidation by a deposit coated tube in a boiler system is described in Squires' work (4). The reaction rate as a function of surface temperature is shown in Figure 2. Surface temperatures in utility boilers range from 800 to 1350 F which covers the active reaction rate temperatures. The SO₃ formed across a given tube in a bank is a function of:

- o partial pressures of SO₂, SO₃ and O₂
- o the reaction rate coefficient
- o Reynolds number (modified for tube pitch/diameter ratio)
- o tube length
- o residence time
- o calibration constant (empirical)

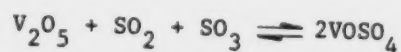
In the model, the boiler is divided into nodes, each node representing a bank of tubes. In the case of this study the test boiler convective pass was broken into 14 nodes. A picture of the convective section of the boiler is

shown in Figure 3. To evaluate the SO_3 formation a plug flow model is used in which the SO_3 at each node is evaluated and accumulated.

The vanadium deposition is assumed to be constant over all surfaces. Reduced vanadium amounts in the fuel are accounted for by reducing the calibration constant proportionally. The residence times are calculated from the geometric input data and the gas mass flow and temperature. The surface temperatures must be input into the program as a function of load. The surface temperatures as a function of generator load for the test boiler are shown in Figure 4. As seen from this figure, the temperatures of the primary superheaters at full load (350 MW) favor a high rate coefficient while the temperatures of the pendant superheaters favors a high conversion rate at low load (93 MW).

Transient SO_3 Formation

Vanadyl sulfate (VOSO_4) can be formed by the interaction of SO_2 , SO_3 , and V_2O_5 by:



The synthesis and decomposition of VOSO_4 has been studied by Weychert and Leyko (5) using a thermogravimetric procedure in a flow system in the temperature range of 680 to 1165 F. The reactants were V_2O_5 and air containing SO_2 in some tests and both SO_2 and SO_3 in other tests. The degree of conversion of V_2O_5 to VOSO_4 was temperature dependent with a maximum conversion occurring at 880 F. The decomposition of VOSO_4 occurred only when a critical temperature, 1080 F, was reached. At 10 to 20 F above this temperature the decomposition was complete. The formation of VOSO_4 in a $\text{VOSO}_4/\text{V}_2\text{O}_5$ mixture as derived from this study is shown in Figure 5. This formation/decomposition curve is used in the model to describe the transient SO_3 formation.

During low load operation the tube surface temperatures are lower and a proportionate amount of VOSO_4 is deposited together with V_2O_5 according to Figure 5. When the load is increased, the tube surface temperatures increase and surfaces which reach temperatures above the critical temperature, liberate SO_3 by the decomposition of VOSO_4 . A finite time is required to decompose all the VOSO_4 .

The generation of SO_3 by the three mechanisms is calculated in the model by the following procedure. The model is run at a low load condition. SO_3 emissions are calculated by homogeneous and catalytic methods. The model is then run a second time at a high load. The convective surface temperatures increase according to temperature vs. load correlation. However, the composition of deposits are kept the same as in the low load case. Surfaces which contain VOSO_4 from the low load case and reach the critical decomposition temperature liberate SO_3 . The model is run a third time at a high load. In this case the compositions of the surfaces are derived from the steady state high load temperatures. Since the compositions are consistent with the temperatures there are no driving forces to decompose the VOSO_4 .

Thus, a low load steady state, high load transient, and high load steady state set of SO_3 emissions can be derived.

TEST PROGRAM

The test series was conducted at a Con Edison facility. The station contains two units which exhaust their flue gas through a single stack. The majority of the testing was performed on a nominal 350 MW unit which includes a twin furnace, face fired B&W boiler. The two furnaces of this boiler, designated as the superheat and reheat furnaces, are identical except the secondary superheater is replaced with a reheater in the reheat furnace. Testing was also performed on the other unit, which is a nominal 500 MW unit including a Combustion Engineering corner fired split furnace with 40 burners. During the program both units utilized a magnesium oxide fuel additive for purposes of keeping the convective passes clean and a magnesium oxide backend additive for air heater corrosion protection. The backend additive was injected as a powder just upstream of the air heaters.

Testing included simultaneous measurements taken at the economizer outlet section (air heater inlet) and the air heater outlet section on the B&W boiler. The measurements included particulate mass by EPA Method 5 and SO_3 by controlled condensation. Boiler operating data, gaseous emission data including CO, NO and O_2 , transmissometer opacity and visual opacity data by EPA Method 9 were taken for each test. A test series included a set of measurements as described above. Three test series were conducted each day for three consecutive days at conditions of: steady state low load, transient, defined as the time when the unit had just reached full load after the morning load ramp, and 3 to 4 hours after full load had been attained. These tests formed the basis for evaluating and correlating the particulate and SO_3 emissions in the model. Particle sizing tests using a University of Washington's Mark 3 cascade impactor were conducted during the first day of testing.

In other tests, measurements were taken simultaneously at the air heater outlet ducts of both boilers. The measurements included the same as indicated above. These measurements formed the basis for correlating the outlet emissions with the stack opacity and comparing the opacity readings with the model predictions.

TEST RESULTS

The SO_3 emissions from the air heater inlet and outlet sections are presented graphically in Figure 6 for one test series. The unit load is superimposed on the graph to display the conditions at which the data were taken. Similarly the particulate emission results for the same series are presented in Figure 7. As can be seen from these figures and from the other test results:

1. The SO_3 emissions at the air heater inlet are higher than at the air heater outlet.

2. The particulate emissions at the air heater outlet are higher than at the air heater inlet. The increase in the particulate emissions in terms of mass loading are approximately equal to the decrease in the SO_3 emissions.
3. The SO_3 emissions at the air heater inlet, on a ppm or lb./Mbtu basis, increase during the transient period and then decrease during steady high load to a lower level than the transient.
4. The particulate emissions at the air heater inlet during the transient period and steady state high load period decrease from the low load condition.

It is evident from these data (refer to Figure 6) that the SO_3 increases during the transient period emanated from the boiler surfaces rather than the air heater surfaces. Also, since the particulate emissions at the air heater inlet do not increase during the transient period, it is concluded that increases in opacity noted during transient periods were not due to a flushing out of stored particulate matter from the boiler surfaces. The increases in the particulate emissions at the air heater outlet were due to a reaction between the SO_3 and magnesium oxide backend additive which occurs either in the air heater itself or, more likely, on the substrate in which the particulate samples were caught.

Visual opacity data were taken throughout the testing periods. The data are plotted against time for a test series in Figure 8. No visual opacity data were taken at low loads since low load operation occurred during nighttime and early morning hours only. The figure reveals high opacities during the transient period compared to the high load steady state condition. Also evident is that the opacity was high for at least 1 hour after full load was reached.

SO_3 MODEL COMPARISON

The B&W boiler data were used to calibrate the SO_3 model in that an a correction factor was applied to the model to force the average SO_3 level to be equal to the measured test data. Comparisons of the measured and predicted SO_3 emissions are presented in Table 1. As can be noted from the table:

1. The homogeneous levels of SO_3 are essentially constant. This is because of a tradeoff between lower furnace temperatures and higher O_2 levels at the lower loads.
2. The catalytic SO_3 levels are higher than the homogeneous levels and increase with load.
3. The decomposition SO_3 levels appear to make up the difference between steady state and transient conditions. In this comparison a theoretical decay rate was assumed for the VO_2 decomposition and the model calculated some residual decomposition SO_3 associated with the steady state high load testing.

OPACITY MODEL COMPARISON

Data from 4 tests were considered applicable for opacity prediction. These tests included simultaneous testing in the outlet ducts of both boilers which exhaust into the common stack. The measurements included SO_3 , particulate mass, transmissometer opacity and visual opacity. The other tests in this study were conducted on one unit only while both units were on line and therefore, were not considered applicable for opacity correlations. The test data were input to RTI's opacity model. The test data were input to RTI's opacity model. Meteorological data were obtained from National Oceanic and Atmosphere Administration summaries for an airport located 9 miles from the Station. The RTI model predicted both in-stack and out-of-stack opacities. Particle size distribution data required for the model were reconstructed from previous test measurements. Table 2 presents the comparisons of the measured and predicted opacities.

CONCLUSIONS

Based on the test results and the model development the following conclusions are made.

1. The test program showed that the generation of SO_3 emissions are primarily responsible for increased opacity levels during both steady state and transient operation on oil firing. Opacity excursions during load increases are due to increases in SO_3 emissions emanating from the boiler surfaces rather than air heater or downstream locations.
2. A model using three mechanisms of SO_3 formation appears to fit the test data. While the homogeneous and catalytic mechanisms are fairly well accepted in the industry, the VOSO_4 decomposition mechanism has not been fully investigated. However, it appears to be a plausible mechanism.
3. A preliminary opacity predictive model had been developed which can be used to predict opacity without requiring emissions test data. The model is written in Fortran for use on an IBM PC. Additional subroutines such as particulate cleaning equipment models can be easily inserted.

FURTHER STUDY

A sensitivity and error analysis must be made on the model. It will include the effects of surface temperatures and emission measurement errors on the calculated particulate and SO_3 emissions, and the stack opacity. This analysis will be performed as part of this project.

Additional field testing to verify the model is required. These should include particulate, SO_3 and opacity measurements on other boilers and on different sulfur content fuels.

Organizations performing fundamental studies on the formation of SO_3 emissions should consider the VO_2 decomposition mechanism.

ACKNOWLEDGMENTS

The work described in this paper was not funded by the U. S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the agency and no official endorsement should be inferred.

The authors would like to acknowledge James B. Homolya, Gunseli Sagun Shareef and George G. Ramsay of Radian Corporation for their efforts in developing the SO_3 formation model.

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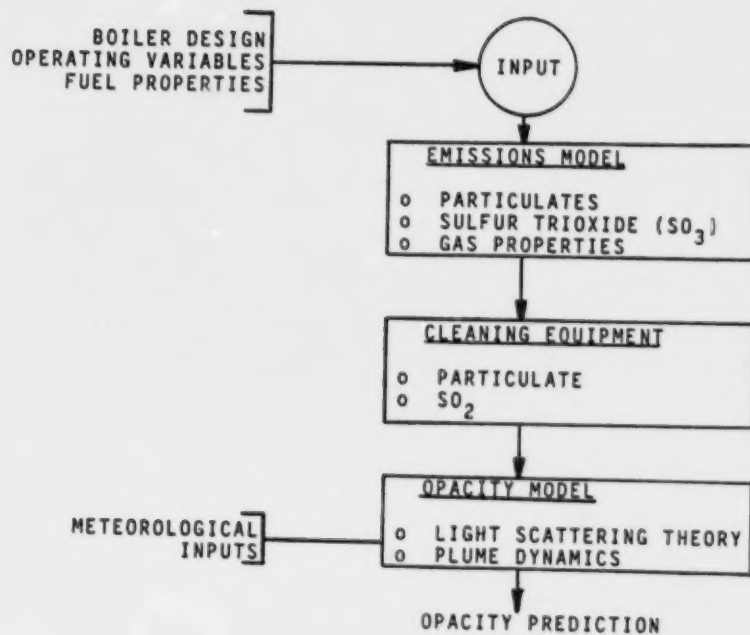


FIGURE 1 MODEL DESIGN

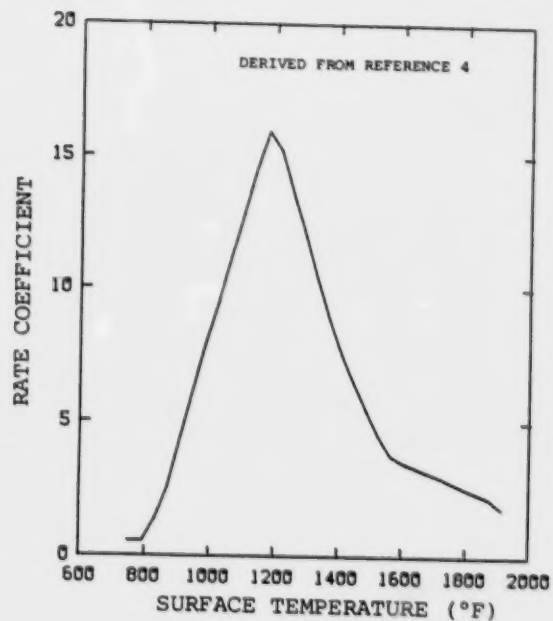


FIGURE 2 REACTION RATE COEFFICIENT FOR CATALYTIC SO_3 AS A FUNCTION OF TEMPERATURE

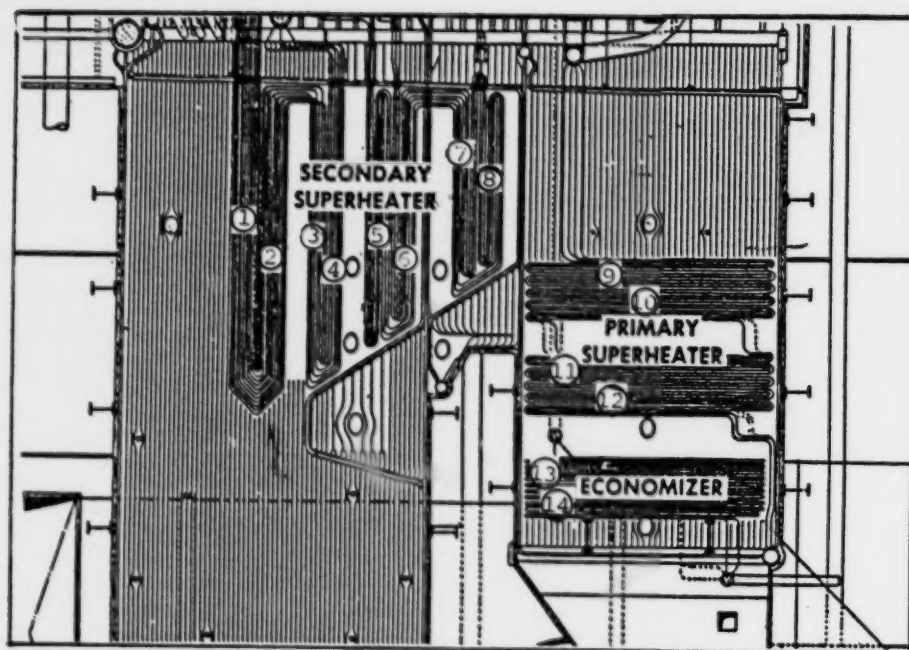


FIGURE 3 CONVECTIVE SECTION OF B&W BOILER

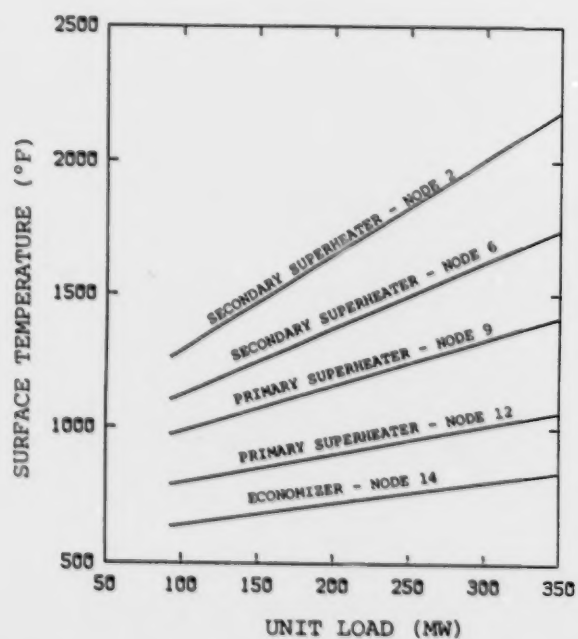


FIGURE 4 HEAT EXCHANGER SURFACE TEMPERATURES AS A FUNCTION OF LOAD

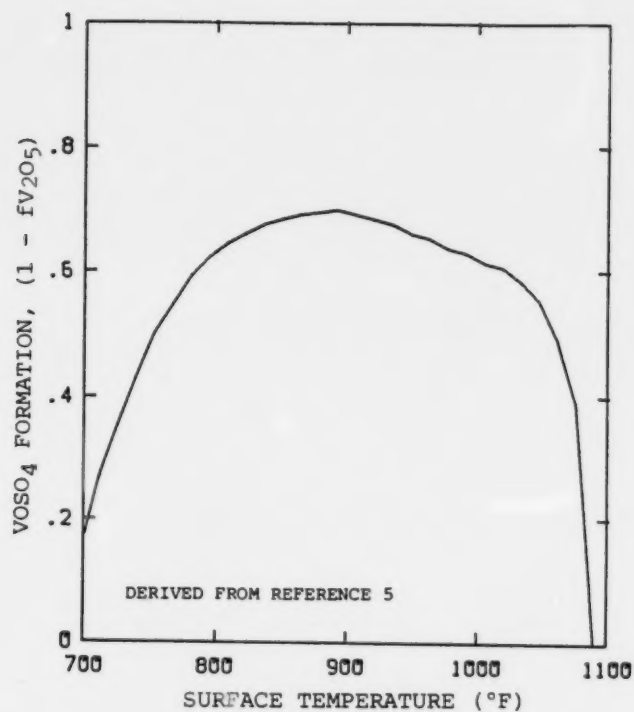


FIGURE 5 VOSO₄ FORMATION AS A FUNCTION OF TEMPERATURE

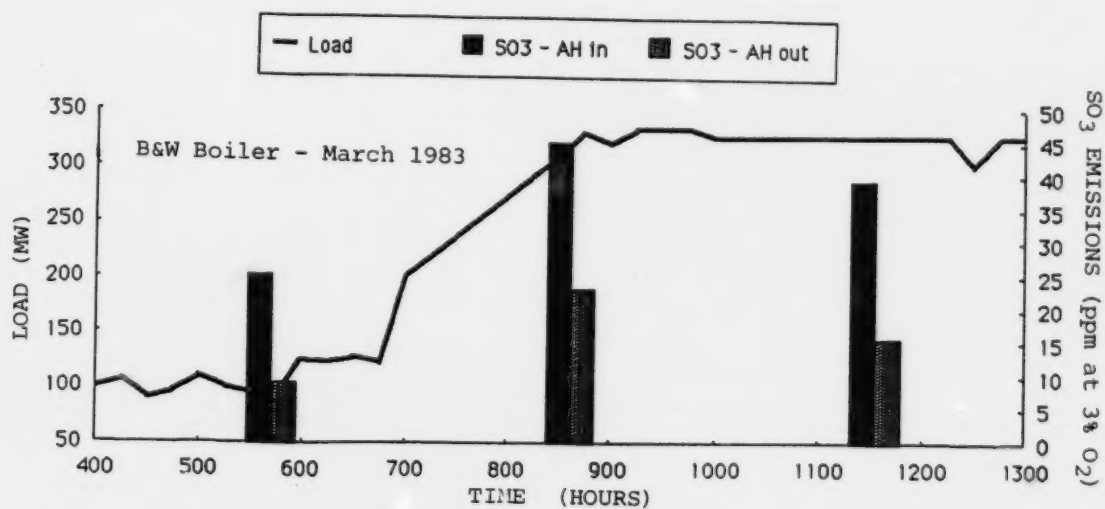


FIGURE 6 SO₃ EMISSIONS FROM AIR HEATER INLET AND AIR HEATER OUTLET

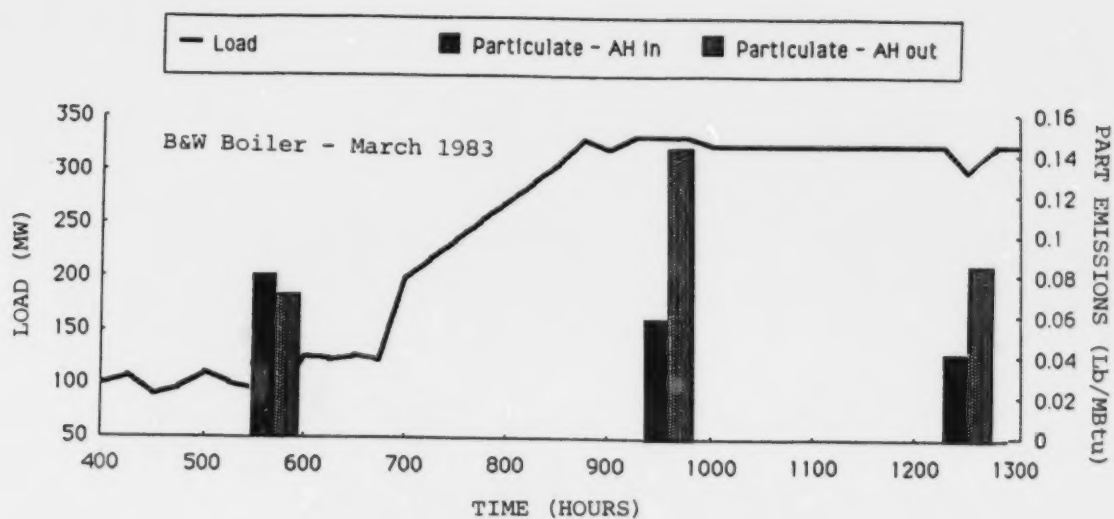


FIGURE 7 PARTICULATE EMISSIONS AT AIR HEATER INLET AND AIR HEATER OUTLET

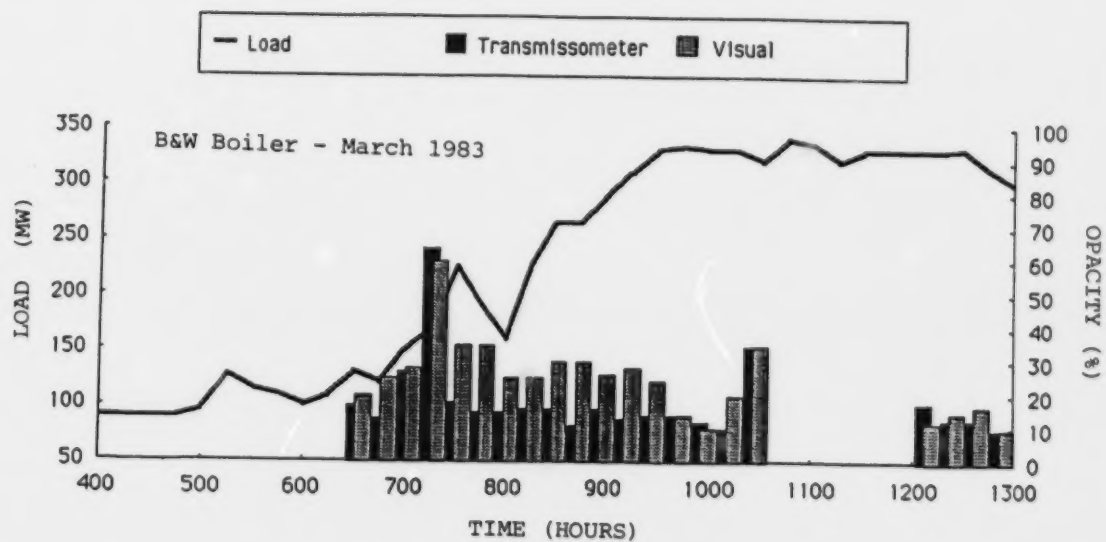


FIGURE 8 VISUAL OPACITY DATA

TABLE 1 COMPARISON OF MEASURED AND PREDICTED SO₃ EMISSIONS

TEST NO.	LOAD (MW)	(SO ₃) _{Hom} (ppm)	(SO ₃) _{Cat} (ppm)	(SO ₃) _{VOSO₄} (ppm)	(SO ₃) _{Pred} (ppm)	(SO ₃) _{Meas} (ppm)	DIFF (P-M) (ppm)
1-1	115	6.2	10.2	0.0	16.4	16.6	-0.2
1-2	326	6.3	14.7	19.5	40.5	36.5	4.0
1-3	320	6.2	14.7	10.1	31.1	31.6	-0.5
2-1	120	6.2	10.3	0.0	16.5	18.2	-1.7
2-2	277	6.1	14.3	18.9	39.3	44.4	-5.1
2-3	335	5.3	12.5	8.4	26.2	32.5	-6.3
3-1	110	6.5	10.7	0.0	17.2	15.1	2.1
3-2	320	6.2	14.6	19.4	40.2	24.3	15.9
3-3	317	6.3	15.0	9.6	30.9	39.1	-8.2

TABLE 2 COMPARISON OF MEASURED AND PREDICTED OPACITIES

TEST NO.	CONDITION	LOAD		MEASURED OPACITY (%)		PREDICTED OPACITY (%)		DIFFERENCE (P-M) (%)	
		(MW)							
		UNIT		IN	OUT OF	IN	OUT OF	IN	OUT OF
		30	20	STACK	STACK	STACK	STACK	STACK	STACK
4-1	STEADY	370	320	14	35	21	42	7	7
5-1	STEADY	150	85	14	45	8	19	-6	-26
5-2	TRANS.	325	285	22	48	12	36	-10	-12
5-3	STEADY	230	285	10	23	8	27	-2	4

THE COLLECTION OF FINE PARTICULATE IN POWER PLANT
ELECTROSTATIC PRECIPITATORS

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ABSTRACT

A case history is presented on the collecting performance on a high efficiency precipitator on thirty trace elements. Data are presented for overall collection efficiencies as well as fractional efficiencies. The results are compared with published fabric filter data.

Measured data show that a precipitator with an overall collecting efficiency of 99.92% is collecting particles in the range 0.02 to 0.2 micron at efficiencies in excess of 99.90%; between 0.2 and 1.0 micron there is a window in performance with a lower collecting limit of 99.1 to 99.5%.

Agglomerate formation plays a significant part in the precipitation/dust collection process and much of the fine particulate is bound up in agglomerate formation which will enhance the precipitation of these fractions.

The thermal history of the trace elements in the boiler combustion process supports the case for compounds of silicon and aluminium providing suitable condensation nuclei for the volatile metals. The data show that higher concentrations of trace elements leaving the precipitator are found in the coarser size ranges and that the particulate concentrations are below the limiting values designated as a health hazard.

INTRODUCTION

Following the initial emphasis in air pollution legislation on the control of particulate emissions to atmosphere, there has been a world-wide interest in the control of the gaseous pollutants which accompany them. For example, current or proposed legislation in the U.S.A., Germany and Japan seeks to limit the quantities of sulphur dioxide emitted by major stationary sources. It seems probable that this approach will be mirrored in other parts of the industrialised world and be extended to cover emission of nitrogen oxides.

An additional area of pollution which is under scrutiny at this time is that of respirable particulate emissions. Amongst the principal sources of such emissions are the metal refining processes in which molten metals are subjected to high temperature processing which result in copious amounts of low micron and sub-micron particulate being formed.

The temperatures to which such processes subject the metals are, however, similar to those which exist within the combustion zones of boiler plant, so that potentially there can be similar fine, respirable particulate issuing from the combustion zone into the gas cleaning plant. It is therefore opportune to examine the emissions from boiler plant and to determine their characteristics relative to any potential public health hazard.

McEvoy and Parker (1) and Lane and Khosla (2) looked at the emissions from both fabric filters and electrostatic precipitators. They were able to demonstrate that the overall mass emissions from well designed electrostatic precipitators was comparable with the mass emissions from a well designed baghouse.

In Figure 1 a comparison is shown between the efficiency of collection of particulate as a function of particle size for three plant. Bag filters are represented by the E.P.R.I. Fabric Filter Pilot Plant (FFPP) at the Arapahoe Test Facility and the Kramer Power Plant Baghouse. The electrostatic precipitator cited is that at the Naughton Power Plant. It is interesting to note the similarity of performance of the precipitator and baghouse. These data were taken from the E.P.R.I. reports of Gooch et al (3) and Ensor et al (4). The E.P.R.I. FFPP data better both in terms of emissions. Since this pilot plant was designed following an operating performance survey of bag filter units throughout industry it is probable that these data represent the optimum performance for a baghouse.

The electrostatic precipitator data which McEvoy and Parker quoted are shown in Table 1. To their data have been added the performance data for the plant chosen to extend the investigation of fine particulate emissions from power plant - the Virginia Electric Power Company (VEPCO), Possum Point Generating Station, Dumfries, Virginia.

The electrostatic precipitator installed on Unit #4 was chosen as a typical state of the art gas cleaning plant operating at high overall mass

collection efficiency, and with a low emission to atmosphere. The typical performance level of the Possum Point plant is shown in the table as 6.4 mg/m³ s.t.p. dry (0.0028 gr/dscf), with all its fields energised. (99.92% collection efficiency).

The VEPCO Possum Point #4 unit precipitator was designed to cope with a 230 MW M.C.R. actual volume of 449 m³/s, 141°C (951,000 acfm at 285°F), at a specific collecting area (SCA) of 106.2 m²/m³/s (539 ft²/1000 acfm). This SCA was based upon only 5 out of the 6 series connected collector fields being energised. Tests run with 5 fields energised have given overall collection efficiencies ranging from 99.81 to 99.87%. The tests on which the present set of fine particulate investigations have been based were for specific collector areas ranging from 140 to 165 m²/m³/s (711 to 838 ft²/1000 acfm).

EXPERIMENTAL

For the purposes of sampling the flue gases, inlet sampling ports were located 9.4 m upstream of the precipitator. At the outlet, sampling was in the stack 45 m above grade level, and 16 m below the stack exit.

Sampling for particulate content was carried out in accordance with EPA method 17. Determination of the mercury content of the flue gas was in accordance with method 101A. Specific particle size analyses at both the inlet and outlet were performed using Andersen stack samplers. The determination of each elemental constituent of the various dust catches was performed using atomic absorption spectroscopy in view of the very small samples involved in the outlet emissions and the stages of the Andersen impactors.

Samples of the coal burnt on Possum Point unit #4 during the tests were taken and both proximate and ultimate analyses were carried out. In addition trace element analyses were carried out.

RESULTS AND DISCUSSION

The data obtained on the test coal samples are shown in Table 2. It can be seen that the ash content was around 10% and that the mean sulphur content, as fired, was 1.30%.

The overall collecting efficiency of the precipitator plant is shown as a function of specific collecting area in Figure 2. The design point was 106.2 m²/m³/s (539 ft²/1000 acfm), 99.70% efficiency. The curve drawn through the data points is typical of the performance degradation as successive fields are de-energised. The efficiency level of 99.54% at an SCA of 105 m²/m³/s (533 ft²/1000 acfm) was achieved with only 4 of the 6 fields energised. The plant can be seen to be functioning satisfactorily within its design emission/efficiency limits.

INFLUENCE OF PARTICLE SIZE

Size dependent chemical analysis over a size range of 0.05 to 10 micron on nearly 30 trace elements in dust entering and leaving a baghouse show the elements to have distribution and collection profiles generally similar to the total mass (5).

The influence of particle size on the efficiency of electrostatic precipitation is shown in Figure 3. In addition to the impactor data obtained on the Possum Point Plant the graph includes the data for a wet electrostatic precipitator and the data from the Naughton Power Plant dry precipitators. Each of the plots shows the characteristic minimum collection efficiency at around 0.2 to 0.8 micron. The impactor data were only capable of resolving down to 0.15 to 0.2 micron particle diameters. The fractional efficiency for dust of particle diameter greater than 12 micron was found to be 99.974% which accounts for the overall dust collection efficiency of 99.92%. These data demonstrate that a well designed precipitator is capable of capturing the sub 10 micron, respirable, dust fraction of the total burden at efficiencies exceeding 99.5%.

The inlet and outlet size distributions of the dust are shown in Figure 4. With 71% w/w of the outlet dust below 10 micron, the emission to atmosphere was 4.5 mg/m^3 (0.002 gr/ft^3). The relatively lower efficiency of precipitation of dust at the fine end of the spectrum can again be seen by the increase of <10 micron fractions in the inlet and outlet from 15 to 71% and for the <1 micron fractions the equivalent increase is from 1.4 to 20%.

INCIDENCE OF TRACE HEAVY METALS

Having so far demonstrated that the fine particulate can be removed at efficiencies in excess of 99.5%, the next question was whether there was a significant rejection of the trace heavy metals such that they were concentrated in the sub-micron size range.

The analyses for the trace metals, including the heavy metals, in the outlet dust impactor samples are listed in Table 3. It can be seen that in every case the concentration of the individual heavy metals was at or below the detection limit of the analytical technique. This limiting accuracy was either 1.3 or $6.3 \text{ } \mu\text{g/m}^3$ with the exception of uranium for which it was $63 \text{ } \mu\text{g/m}^3$.

The level of mercury at both the precipitator inlet and outlet was found to be below the analytical minimum of $22 \text{ } \mu\text{g/m}^3$. Further investigations at plant where the mercury level in the coal is higher will be required to enable determinations of removal efficiency to be made.

Parker and McEvoy demonstrated that on the Agecroft Power Plant in the U.K. the highest concentrations of the heavy metal elements was in the fraction greater than 8.6 micron. It is interesting to note that with the

Possum Point data this trend is exhibited by the zinc content of the emissions. There is no strong trend in the Possum Point data to indicate rejection of the finer size fractions of the dust.

The fact that there is no significant increase in trace heavy metal content of the fine fraction end of the spectrum lends support to the theory proposed by McEvoy and Parker viz the condensing sub-micron metal particulate attaches itself to the larger silica/silicate particles producing the characteristic overlay of condensed particles on their surface.

There is a typical example of this condensate in Figure 5 where an approximately 100 micron diameter particle can be seen to be literally covered in sub-micron material condensed onto the surface as the particle progressed from the combustion zone through to the inlet of the precipitator from which it was sampled.

The ability of electrostatic precipitators to capture low micron and sub-micron material should not come as any surprise since there are plant collecting metallurgical fume in a variety of different smelting processes. Some examples of these were given by McEvoy and Parker and are reproduced in Table 4, together with some additional examples. Although detailed particle size analysis data are not available for all these fine fumes, the limited data which we have indicates that 60 to 100% of the dust particles at the inlet to the gas cleaning units are below one micron in diameter.

TRACE HEAVY METAL COLLECTION EFFICIENCIES

Data previously obtained (1) in an analysis of dusts collected isokinetically at the inlet and outlet of the Dave Johnston Steam Plant have been used to quantify the specific elemental collection efficiencies for each of a range of trace heavy metals. These data are summarised in Table 6.

The quantities of each element emitted to atmosphere correspond closely in each case to their being removed at essentially the same efficiency as the overall particulate removal efficiency of 99.975%. Determination of the mercury and fluoride contents of the outlet emissions showed them to be below $0.23 \mu\text{g}/\text{m}^3$ which is consistent with collection efficiencies in excess of 99.975%.

For all the elements listed in Table 6 the total emission to atmosphere amounted to less than $0.01 \text{ mg}/\text{m}^3$. This is 0.46% of the $2.3 \text{ mg}/\text{m}^3$ ($0.001 \text{ gr}/\text{ft}^3$) total particulate emission. Calculated on an annual emission basis, some 80 kg (177 lb) of these trace heavy metals are emitted from the stack.

To put these concentrations into context it is interesting to compare them with the data produced by Mansi et al (6). They recorded the ground level concentrations of heavy metals in the Naples area of Italy. Their

data, they concluded, were similar to those obtained by others in the Urban areas of various European and American cities. A comparison is made in Table 5 between their data and the stack emissions at Dave Johnston Steam Plant. It can be seen that with the exception of manganese the Power Plant data fall below the Italian ground concentration levels.

It is clear that with dispersion of the stack plume, the resultant ground level concentration of the various elements will fall to well below the prevailing background levels found in Urban areas (7). Even at the stack outlet the concentration of the trace heavy metals falls significantly short of the threshold limit values (8, 9).

As previously stated, McEvoy and Parker were also able to show that the elemental concentrations of trace heavy metals were not at their highest levels in the <2 micron fraction of dusts sampled at the Agecroft Power Plant of the Central Electricity Generating Board in the U.K.. Indeed, by far the highest concentration of each element analysed was to be found in the +8.6 micron fraction with 20 to 80 times the <2 micron fraction content of the trace heavy metals found in the +8 micron fraction of the total outlet dust samples.

CONCLUDING REMARKS

The data which have been analysed and presented herein show that well designed and properly maintained electrostatic precipitators are capable of capturing fine particulate and the associated trace heavy metal elements at high efficiencies. Even though the characteristic "window" of penetration of particulate in the size range 0.2 to 1.0 micron leads to reduced capture in that range, the data show that the individual trace elements are retained by the precipitator at levels of efficiency similar to those of the elements which make up the bulk of the boiler fly ash.

The heavy metal trace elements are found in concentrations which are significantly less than their threshold limit values at the stack outlet. Dilution by natural dispersion indicates that at ground level they should not represent a significant health hazard. End users must, however, always make their own determination of the potential hazards in the light of proposed environmental and operating conditions, and of the latest updated environmental health statutes and regulations.

The authors recognise that some of the elements for which particulate analysis are reported herein may be present in a condensible form not collected as particulate at the flue gas sampling temperature. To ensure capture of any highly volatile heavy metal compounds would require significant reduction of the flue gas temperature. It is therefore possible that the total mass emission of such compounds could lead to trace heavy metal emissions which exceed the threshold limit values.

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TABLE 1. PRECIPITATOR PARTICULATE EMISSION DATA

Plant	Emission	
	(gr/ft ³)	(mg/m ³)
Centralia #1,2	0.0002	0.5
Naughton #1,2	0.0003	0.7
La Cygne #2	0.0032	7.3
Dave Johnston #1,2,3	0.0017	3.9
George Neal #3	0.0070	16.0
Iatan #1	0.0050	11.5
Hawthorn #1,2	0.0056	12.8
Northeastern #3,4	0.0064	14.7
Boardman #1	0.0014	3.2
G.R.D.A. #1	0.0089	20.4
Possum Point #4	0.0028	6.4

TABLE 2. POSSUM POINT MEAN COAL PROPERTIES DURING TESTS

(Per cent w/w)	
Ash	10.0
Carbon	74.73
Hydrogen	5.04
Nitrogen	1.42
Oxygen	7.51
Fixed Carbon	54.55
Sulphur	1.30
Moisture	9.26
Volatile Matter	35.45
Heat Value	31,157 kJ/kg
	13,395 Btu/lb

TABLE 3. OUTLET SAMPLE ELEMENTAL ANALYSES
POSSUM POINT POWER STATION

Size Range (micron)	+8	6-8	4-6	2-4	1-2	0.5-1	-0.5
Emission in Range (Per cent)	35	10	12	15	8	8	12
Element	Concentration ($\mu\text{g}/\text{m}^3$)						
Aluminium	2062	138	238	206	231	375	687
Antimony	<1.3	<1.3	<1.3	<1.3	<1.3	<1.3	<1.3
Arsenic	<6.3	<6.3	<6.3	<6.3	<6.3	<6.3	<6.3
Barium	69	3.0	2.6	4.2	3.6	5.0	6.9
Boron	88	69	88	69	94	131	231
Cadmium	<1.3	<1.3	<1.3	<1.3	<1.3	<1.3	<1.3
Calcium	387	306	456	344	512	812	1187
Chromium	<1.3	<1.3	<1.3	<1.3	<1.3	<1.3	<1.3
Cobalt	<1.3	<1.3	<1.3	<1.3	<1.3	<1.3	<1.3
Copper	<1.3	<1.3	<1.3	<1.3	<1.3	<1.3	<1.3
Iron	46	22	25	21	23	23	35
Lead	<6.3	<6.3	<6.3	<6.3	<6.3	<6.3	<6.3
Magnesium	75	58	88	75	100	144	237
Manganese	<1.3	<1.3	<1.3	<1.3	<1.3	<1.3	<1.3
Molybdenum	<1.3	<1.3	<1.3	<1.3	<1.3	<1.3	<1.3
Nickel	2.0	<1.3	<1.3	<1.3	<1.3	<1.3	<1.3
Potassium	9.4	7.5	9.4	7.5	10	9.4	10.6
Selenium	<6.3	<6.3	<6.3	<6.3	<6.3	<6.3	<6.3
Silicon	138	113	163	156	188	275	412
Silver	<1.3	<1.3	<1.3	<1.3	<1.3	<1.3	<1.3
Sodium	100	106	100	88	106	100	150
Titanium	88	6.2	9.4	8.7	8.7	14	7.5
Uranium	<63	<63	<63	<63	<63	<63	<63
Vanadium	<1.3	<1.3	<1.3	<1.4	<1.3	<1.3	<1.3
Zinc	3.1	1.8	2.6	1.6	1.6	1.6	1.3

Condensable and Particulate Mercury was determined at less than $22 \mu\text{g}/\text{m}^3$

TABLE 4. TYPICAL APPLICATIONS OF ELECTROSTATIC PRECIPITATORS
IN THE CAPTURE OF METALLURGICAL FUME

Fume Composition	S.C.A.		Efficiency (Per Cent)	Temperature		Emission	
	(ft ³ /1000 acfm)	(m ³ /m ³ /s)		(°F)	(°C)	(gr/ft ³)	(mg/m ³)
Chromium Oxide	337	66.9	99.6	392	200	0.0079	18
Chromium Sulphate	324	64.4	96.7	199	93	0.0009	2
Arsenic, 92% As ₂ O ₃	396	78.7	96.0	230	110	0.0332	76
Manganese, 52% MnO ₂	250	49.6	91.0*	550	288	0.4034	923
Copper, 30% CuO	679	134.8	99.94	559	315	0.0039	9
Molybdenum, 58% MoS ₂	421	83.6	98.5	649	343	0.0249	57
Iron Oxide	535	106.3	99.85	351	177	0.0013	3
Gold	209	41.5	99.58	176	80	0.0017	4
Lead	374	74.2	98.67	126	52	0.0013	3

* As a Pre-cleaning System

TABLE 5. COMPARISON BETWEEN INDUSTRIAL/URBAN HEAVY METAL
GROUND LEVEL CONCENTRATIONS AND DAVE JOHNSTON POWER PLANT
STACK EMISSIONS

Element	Particulate Concentrations (µg/m ³)		Stack Emission (µg/m ³)
	Industrial	Urban	
Chromium	0.06 - 0.19	0.05 - 0.52	0.321
Copper	0.25 - 1.45	0.11 - 2.68	1.168
Lead	0.52 - 4.52	1.05 - 5.12	0.344
Manganese	0.09 - 0.43	0.07 - 0.62	1.832
Nickel	0.05 - 0.24	0.02 - 1.69	0.527

TABLE 6. TRACE HEAVY METAL CONCENTRATION DATA -
DAVE JOHNSTON STEAM PLANT

Element	Concentration ($\mu\text{g}/\text{m}^3$)		Collection Efficiency (Per Cent)
	Inlet	Outlet	
Antimony		<0.115	
Arsenic		<0.115	
Beryllium		<0.115	
Boron	2290.0	0.527	99.977
Cadmium	91.6	0.046	99.950
Chromium	1374.0	0.321	99.977
Cobalt	274.8	0.092	99.967
Copper	4580.0	1.168	99.975
Lead	824.4	0.344	99.958
Manganese	6228.8	1.832	99.971
Molybdenum		<0.115	
Nickel	458.0	0.527	99.885
Selenium		<0.115	
Silver	366.4	0.137	99.963
Strontium	17404.0	3.435	99.980
Thorium	1557.2	0.366	99.977
Zinc	2198.4	0.756	99.966

Overall Particulate Collection Efficiency during Tests:-
99.975%, Outlet Emission $2.3 \text{ mg}/\text{m}^3$
 $0.0010 \text{ gr}/\text{ft}^3$

FIGURE 1

PENETRATION/ EFFICIENCY
AS A FUNCTION OF PARTICLE SIZE

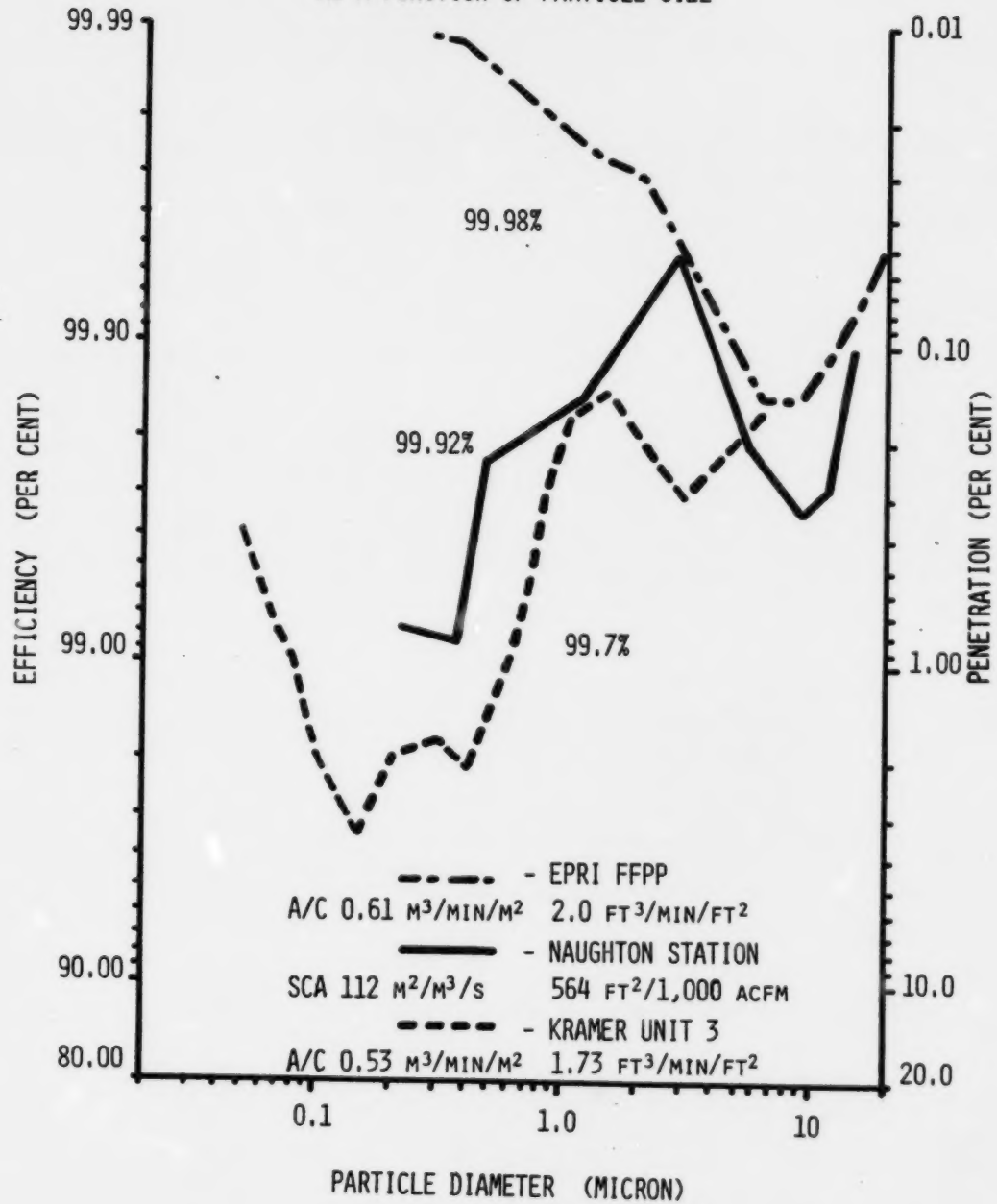


FIGURE 2

POSSUM POINT POWER STATION
PENETRATION/EFFICIENCY AS A FUNCTION
OF SPECIFIC COLLECTING AREA
(s.c.a.)

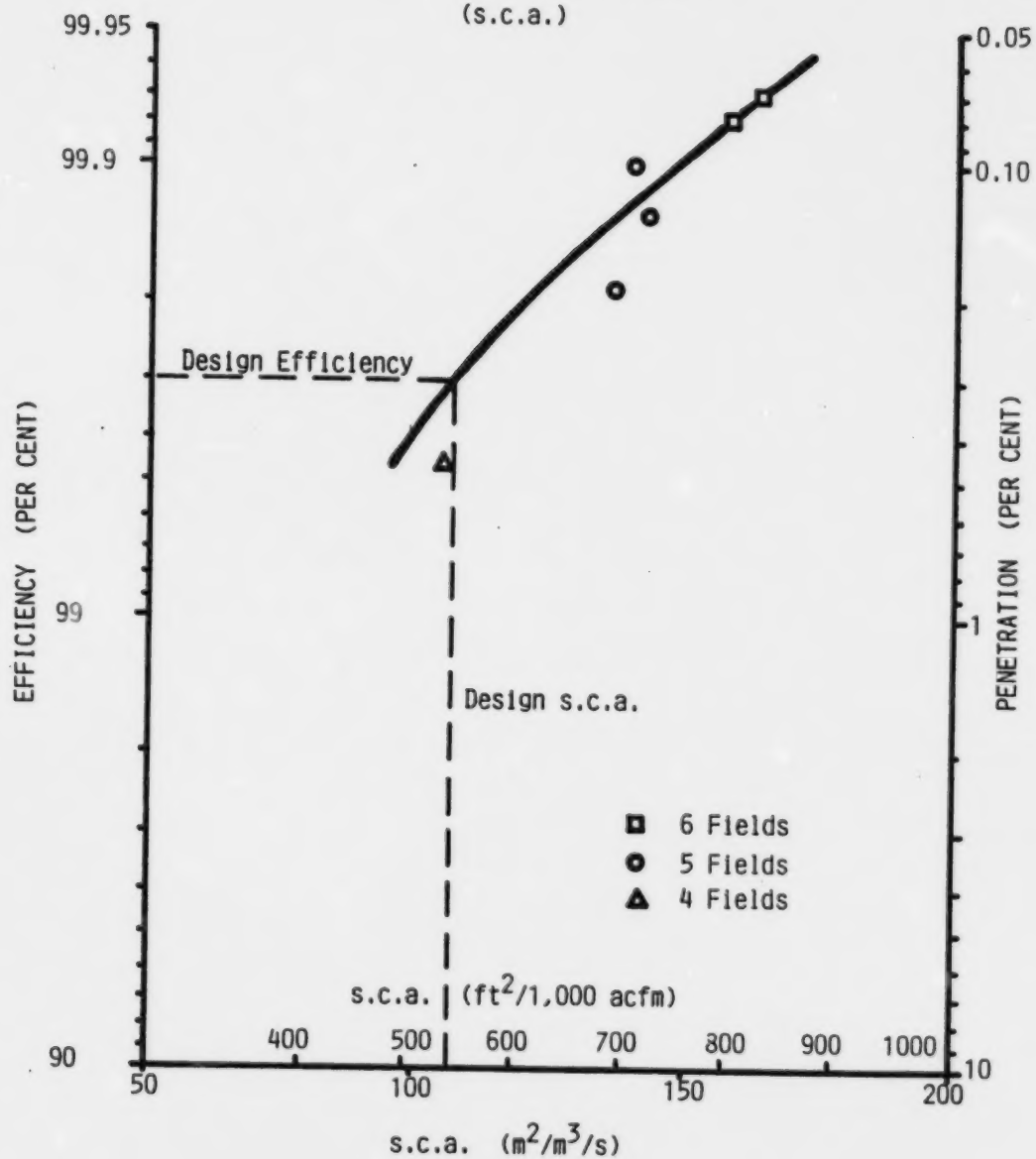


FIGURE 3

EFFICIENCY/PENETRATION AS A FUNCTION
OF PARTICLE SIZE

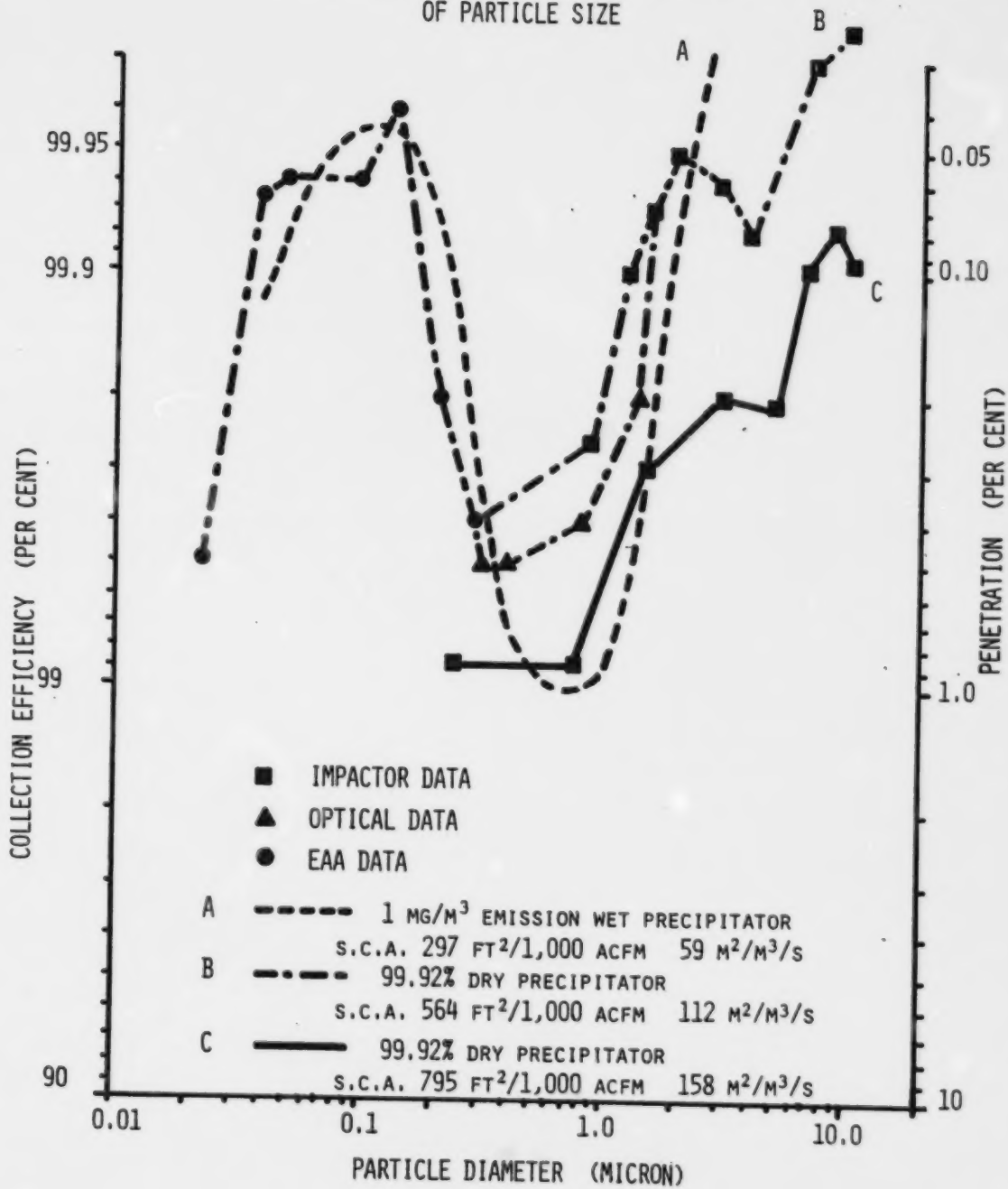


FIGURE 4

PARTICLE SIZE ANALYSES

POSSUM POINT DUST

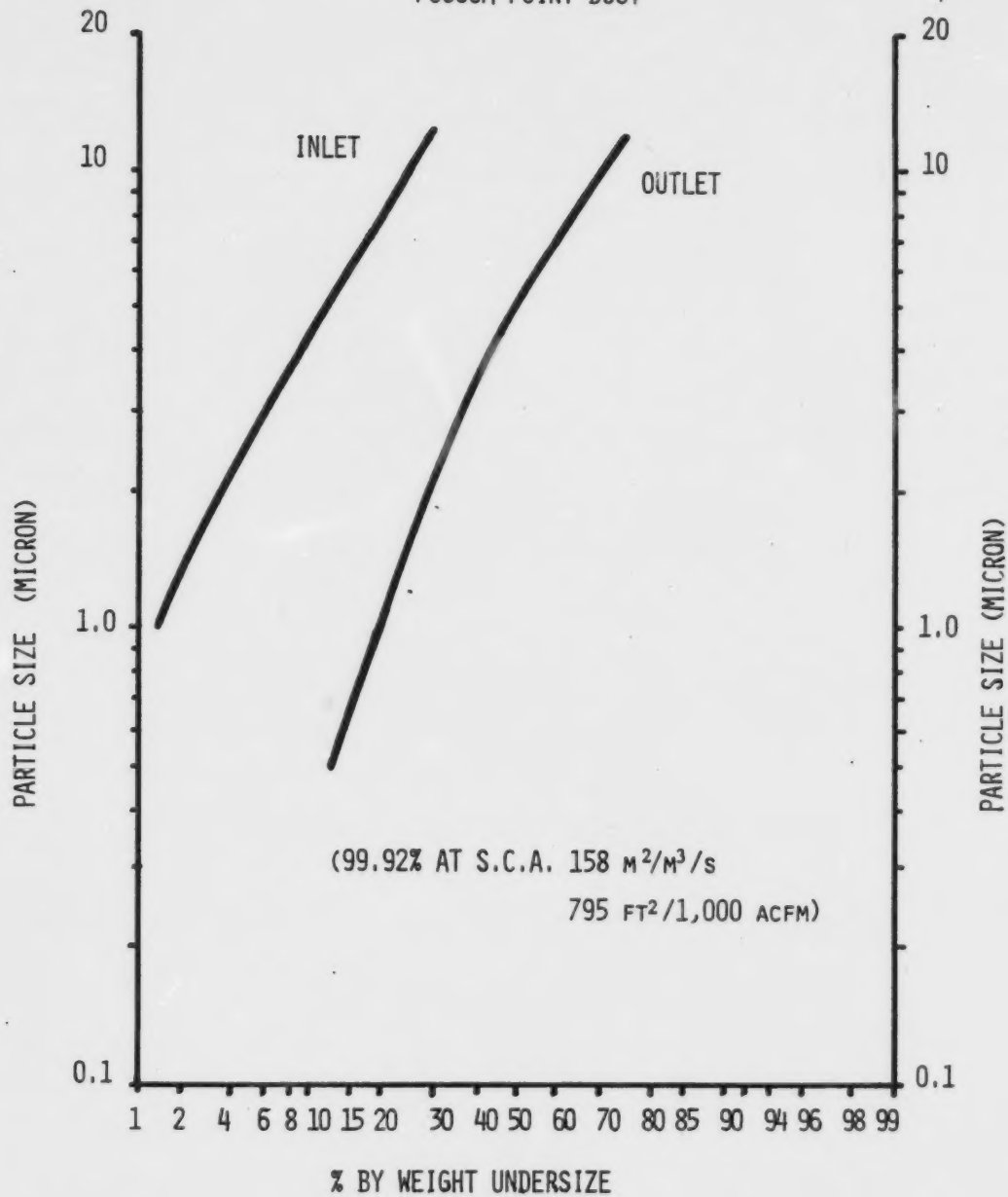
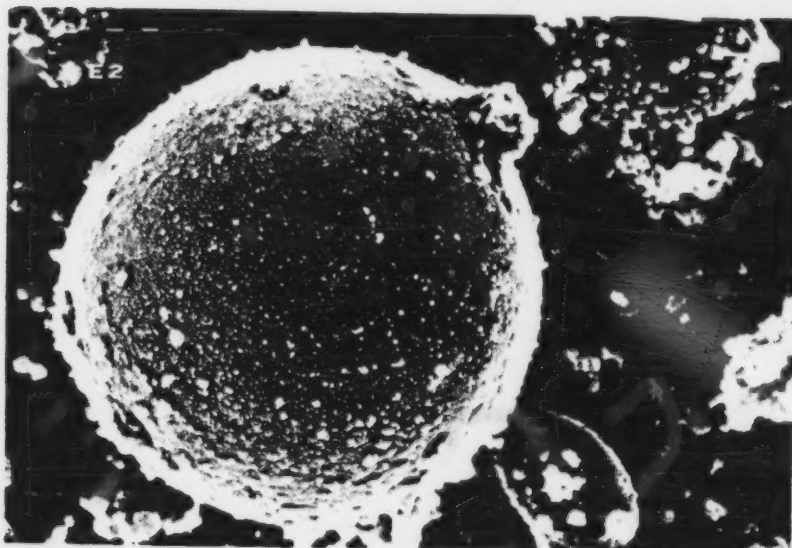


FIGURE 5
100 MICRON DUST PARTICLE EXHIBITING
THE OVERLAY OF SUBMICRON PARTICLES



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PREDICTING THE PARTICLE SIZE DISTRIBUTION OF FLY ASH

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ABSTRACT

This paper describes the development and application of a set of mathematical models to predict the size distribution of fly ash particles formed in pulverized coal-fired utility boilers. The formation of large ($>0.2\text{-}\mu\text{m}$) fly ash particles is described in terms of a simple breakup model based on complete coalescence of molten mineral inclusions within fragments of the devolatilized coal char. A volatilization/nucleation model and a coagulation model are used to treat the formation of ultrafine ($<0.2\text{-}\mu\text{m}$) particles.

The size distributions predicted by the models are compared to particle size data obtained from a pilot-scale combustor and a full-scale utility boiler using an electrical aerosol size analyzer, an optical counter, and cascade impactors. The effects of coal ash content, ash composition, coal particle size, and flame temperature on the predicted size distribution are examined, and the implications on control device performance are discussed.

INTRODUCTION

When a new coal-fired power plant is being designed, the particle size distribution of the fly ash must be known in order to properly size an electrostatic precipitator for control of particulate emissions. The set of mathematical models described in this paper provides a method of predicting the fly ash particle size distribution from the fundamental parameters that are available in the plant design stage. These parameters include the properties of the coal and the design conditions of the boiler that influence the fly ash formation mechanisms. By employing a fundamental treatment of the physical and chemical mechanisms involved in ash particle formation, this model is able to provide physical insights that are not possible with empirical approaches. For example, the treatment of ash volatilization may be useful in studies of boiler tube fouling, which is initiated by deposition

of volatilized ash components (1,2). Based on comparisons with particle size measurements, the mechanistic treatment also appears to provide an accurate predictive tool suitable for use in a precipitator design procedure.

MINERAL MATTER IN COAL AND ITS RELATION TO FLY ASH

The mineral matter in coal occurs in three physical modes: inherent, included, and extraneous (3). Inherent mineral matter refers to the various minerals that are organically attached to the coal matrix. The inherent mineral matter originated from the inorganic components of the plant cells from which the coal was formed, although some of the original inorganic species may have been replaced by an ion exchange process with percolating water during the coalification sequence (4). The included mineral matter is present in the form of discrete, microscopic islands of minerals that were trapped in the coal matrix during coalification (5). Most of the mineral matter (typically 60 to 80%) is present in the included form. Extraneous mineral matter may account for up to 20% of the total mineral content. Extraneous mineral matter is simply soil and rock that is unintentionally mined along with the coal. It is easily removed in coal preparation plants.

CHARACTERISTICS OF MINERAL MATTER RELATIVE TO COMBUSTION

The three forms of mineral matter behave differently under combustion conditions. A given mineral may also exhibit a different type of behavior in intimate contact with a coal particle than it would as a separate entity suspended in the oxidizing furnace gases. For example, as a result of the intimate bonding between the inherent mineral matter and the carbonaceous matrix, the inherent mineral matter is volatilized to a greater extent than the other mineral forms. Any extraneous mineral matter that is not removed by coal preparation or by the classifying action of the coal pulverizers will generally be larger in particle size than the other mineral forms. It will, therefore, tend to drop out as bottom ash and not contribute to fly ash formation. The included mineral matter is released from the coal through a series of steps involving the fragmentation of burning char particles, burnout of the char fragments, and coalescence of the molten mineral inclusions as they follow the receding surface of the char (6).

RELATIONSHIP BETWEEN MINERAL INCLUSIONS AND FLY ASH PARTICLE SIZE

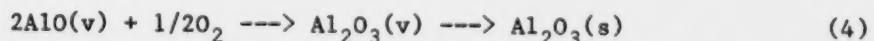
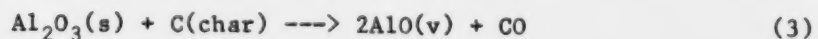
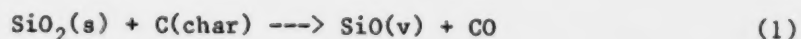
Optical microscopy of coal thin sections has shown that the mineral inclusions typically have a mass median size of about 2 μm (7). This is an interesting observation because the fly ash from pulverized coal combustion exhibits a bimodal size distribution, with one mode typically centered about a particle size of 0.1 to 0.2 μm and the other mode centered about 10 to 20 μm (8,9,10). On a mass basis, the supermicron mode is predominant, generally accounting for over 90% of the particulate mass. On a number basis, however, the submicron mode dominates, accounting for over 90% of the particles. Thus, fly ash consists of a relatively large number of small ($\sim 0.1\text{-}\mu\text{m}$) particles and a relatively small number of large ($\sim 10\text{-}\mu\text{m}$) particles, while the mineral inclusions are of an intermediate size on the order of 1 μm . Two types of mechanisms must be invoked to explain this observation: a mechanism whereby a

relatively small number of large ash particles are formed from mineral inclusions that are an order of magnitude smaller, and a mechanism whereby a relatively large number of small ash particles are formed from mineral inclusions that are an order of magnitude larger. A breakup/coalescence mechanism (7,11) has been advanced for the former. Two different mechanisms have been proposed for the latter: one involving the volatilization and homogeneous nucleation of ash components (7,11,12) and one involving the disintegration of molten ash droplets by the bursting of gas bubbles (13,14,15).

BEHAVIOR OF COAL MINERAL MATTER UNDER COMBUSTION CONDITIONS

When pulverized coal particles are injected into the flame zone through the burner, they are subjected to intense heating by radiation and conduction/convection heat transfer from the hot combustion gases. The heating rates have been estimated to be on the order of 100,000 to 1,000,000 °K/sec (16). Under these conditions, the volatile organic components of the coal are rapidly vaporized and transported into the bulk gas, where they ignite and burn to form CO₂ and H₂O together with small amounts of SO₂, SO₃, and NO_x. The devolatilization process leaves behind carbonaceous residues, or char particles, characterized by considerable internal pore structure, cracks, and fissures. This is illustrated in Figure 1. At this point, the mineral inclusions are still trapped within the char particle.

The next stage in the combustion process is char oxidation or carbon burnout. In this stage, the oxygen-consuming reactions occurring at the char particle surface cause an oxygen concentration gradient to be established within a thin layer of gas surrounding the particle. This is illustrated in Figure 2. The rate of combustion is controlled by the rate of diffusion of oxygen through this thin gas layer. Consequently, reducing conditions exist at the char particle surface even though oxidizing conditions prevail in the bulk gas. This is an important point because under these conditions certain oxides in the mineral matter can be reduced to volatile suboxides. The volatile suboxides can then be transported into the gas phase, oxidized, and condensed as the stable oxides. This is a major mechanism of ultrafine particle formation in the coal combustion aerosol, and it explains the occurrence of refractory oxides in submicron fly ash particles. The following reactions illustrate this concept as applied to two of the most common refractory oxides, silica and alumina:



COALESCENCE OF MOLTEN MINERAL INCLUSIONS AND FRAGMENTATION OF CHAR

As the char is burned away, the char particle surface recedes, exposing the mineral inclusions. The minerals remain attached to the char particle surface due to their high surface tension, which has been estimated to be about 320 dynes/cm (17). As the char particle shrinks, the minerals on the surface come into contact and coalesce. If the char particle does not fracture during this process, all of the mineral inclusions would be fused into a single ash particle upon complete burnout of the char. Laboratory combustion experiments have shown, however, that three to five ash particles are formed from each char particle (7). Therefore, the char particle must break up into several fragments during the burnout process. The breakup presumably occurs along the cracks and fissures formed during devolatilization. A mathematical description of this process, called the breakup model (11), has been used to explain the formation of the large particle mode of fly ash.

VOLATILIZATION AND NUCLEATION OF ASH COMPONENTS

The volatile species in coal ash include several metals with low boiling points (e.g., Hg, Cd, Se, As, and Zn). Due to their volatility, these metals are found to be concentrated in fine particles and on particle surfaces (18, 19). While their concentrations are enriched in the ultrafine particulate, these metals still account for only a relatively small amount of the submicron ash. In terms of its major components, the composition of the submicron ash is actually quite similar to that of the supermicron ash (14,15). Silica and alumina typically account for 80% of the submicron particulate mass. Therefore, reduction and volatilization of the refractory oxides, rather than vaporization of low boiling metals, play the dominant role in forming the submicron fly ash particles. The volatilization of these components has been observed in laboratory furnace experiments (7,20,21). Mims et al. (20) measured the volatilization rates of various elements from coal mineral matter. Based on an analysis of their rate data, they showed that the volatilization rates were controlled by diffusion of the volatile species away from the particle surface. A mathematical treatment of their results may be used to predict the amount of a given refractory oxide that will be volatilized under prescribed conditions.

As the volatile suboxide diffuses away from the char particle and is exposed to increasing oxygen concentrations, it is converted to the stable oxide and rapidly nucleates. McNallen et al. (22) calculated that homogeneous nucleation of silica can take place even in the presence of a large concentration of preexisting particles, provided that a sufficient degree of supersaturation is attained. This condition is easily satisfied because the partial pressure of the volatile suboxide is much larger than the vapor pressure of the stable oxide. The size and concentration of the nuclei that are formed in this process may be predicted by homogeneous nucleation theory. The application of this concept to the formation of submicron fly ash has been discussed in the literature (23,24). Prior to the present study, however, there was no comprehensive mathematical description of the volatilization/nucleation mechanism of ash formation.

CENOSPHERE FORMATION AND BUBBLE BURSTING

The dehydration and decomposition of certain minerals can cause the formation of gas bubbles within molten ash droplets. If the ash particle solidifies before the gas can escape, the result is a hollow ash particle called a cenosphere. Under the electron microscope, cenospheres are evident in most ash samples; however, they do not account for a significant portion of the total ash. Evidently, most of the gases escape prior to ash solidification, and the bubbles collapse leaving a nonporous ash particle. If the gases do not escape, the pressure within the bubble can become high enough to burst the molten ash droplet. Ramsden (13) offered this as an alternative to the volatilization/nucleation model of ultrafine ash particle formation. Bubble bursting generally tends to produce particles with a broad size distribution, whereas nucleation processes tend to produce a relatively narrow range of particle sizes. In situ particle size measurements at coal-fired power plants have shown that the ultrafine mode is a very narrow spike. This strongly suggests that the volatilization/nucleation mechanism is predominant over the bubble bursting mechanism in the formation of ultrafine fly ash particles. For these reasons, only the volatilization/nucleation mechanism is considered in the present study to account for the submicron size fraction.

DESCRIPTION OF THE MATHEMATICAL MODELS

The foregoing concepts were used to develop three mathematical models that can be used in combination to predict the size distribution of fly ash formed in pulverized coal-fired utility boilers. Two FORTRAN programs and one BASIC program were written to implement the three mathematical models. The first program, BREAK.FOR, is based on the breakup model discussed earlier and is used to predict the size distribution of the large fly ash particles. The second program, NUCL.FOR, is based on the volatilization/nucleation model of nuclei formation and is used to predict the size of nuclei that are condensed from the volatilized ash components. The third program, HTOAG.BAS, models the formation of the ultrafine mode by coagulation of the nuclei predicted by NUCL.FOR.

THE BREAKUP MODEL OF LARGE PARTICLE FORMATION (BREAK.FOR)

The mathematical development of the breakup model is based on a simple ash mass balance in which each coal particle releases p ash particles. From the mass balance, it follows that

$$d_a = d_c (x\rho_c / p\rho_a)^{1/3} \quad (5)$$

where d_a is the diameter of an ash particle formed from a coal particle of diameter d_c , x is the mass fraction of ash in the coal, and ρ_c and ρ_a are the densities of the coal and ash. Equation (5) can be used to predict the size of the ash particles that will be formed from a given coal particle size.

To relate the mass distributions of the coal and ash particles, the particle diameters d_c and d_a may be viewed as the midpoints of vanishingly

small size intervals that contain mass increments of dM_c and dM_a . If the ash fraction x does not vary with particle size, then

$$dM_a = x dM_c \quad (6)$$

From Equations (5) and (6), it may be shown that

$$dM_a/dd_a = x(\rho_a/\rho_c)^{1/3} (dM_c/dd_c) \quad (7)$$

which gives the relationship between the differential mass distributions of the ash and coal. The differential mass distribution of the coal has been found to fit a special form of the Rosin-Rammler distribution (25,26):

$$dM_c/dd_c = (1.2M_c/d_c)(d_c/\bar{d}_c)^{1.2} \exp[-(d_c/\bar{d}_c)^{1.2}] \quad (8)$$

where $\bar{d}_c$ is the mean coal particle size. If Equation (8) is substituted into Equation (7), M_c is replaced by M_a/x , and both sides are divided by M_a , the result is

$$(1/M_a)(dM_a/dd_a) = (1.2/d_c)(d_c/\bar{d}_c)^{1.2} (\rho_a/\rho_c)^{1/3} \exp[-(d_c/\bar{d}_c)^{1.2}] \quad (9)$$

which gives the normalized mass distribution of the fly ash particles. A logarithmic distribution of the form $dM_a/d\ln d_a$ is often used in the literature and can be obtained by multiplying both sides of Equation (9) by d_a . After combining this with Equation (5), the result is

$$(1/M_a)(dM_a/d\ln d_a) = 1.2(d_c/\bar{d}_c)^{1.2} \exp[-(d_c/\bar{d}_c)^{1.2}] \quad (10)$$

which gives the normalized differential mass associated with the ash particle size given by Equation (5). The normalized size distribution for the large particle mode is constructed by applying Equations (5) and (10) over the entire range of coal particle sizes and plotting the results. This may be converted to a non-normalized form by multiplying by the total mass loading.

The lower limit of the large particle mode is determined by the lower limit of the coal particle size. For practical purposes, this limits the applicability of the breakup model to ash particles larger than about 0.2 μm .

THE VOLATILIZATION/NUCLEATION MODEL (NUCL.FOR)

In this model, the gas-phase concentrations of the volatile suboxides are predicted by fitting the volatilization rate data of Mims et al. (20) to a gas-phase diffusion model and using the thermodynamic predictions of Ulrich et al. (23). In the model, volatilization is assumed to occur only during char burnout, with the char particle modeled as a shrinking sphere. The total amount of material volatilized is determined by integrating the volatilization rate versus time up to the point of complete burnout. The time required for complete burnout is predicted from a shrinking core model of the burning char. The total amount of material volatilized is expressed as a partial pressure and compared to published vapor pressure data on the stable oxides. This ratio defines the degree of supersaturation for each volatilized component.

The mixing rules of Kiang et al. (27) are then applied to determine the overall degree of supersaturation of the multicomponent system. The nucleation of the mixture is then treated by classical homogeneous nucleation theory. The size of the nuclei is calculated by the Kelvin equation, and the nucleation rate is calculated by the classical rate equation of Volmer and Weber (28). These equations are coupled to the volatilization rate equations and applied over the entire time-temperature profile of the boiler until all of the volatilized material has been nucleated. The number concentration of nuclei is computed by integrating the nucleation rate versus time.

The results reported in this paper were obtained with a simplified version of this model in which only two components, silica and alumina, were considered. This was done to allow the use of binary mixing rules and avoid the complex mathematics of multicomponent nucleation calculations. Since silica and alumina typically account for 80% of the submicron aerosol mass, the error introduced by this simplification should be about 20%. In view of the accuracy of ultrafine particle size analyzers, this error is considered to be acceptable.

The mathematical details of the volatilization/nucleation model are beyond the scope of this paper. A complete derivation and a listing of the computer code may be found in reference (29).

THE COAGULATION MODEL OF SMALL PARTICLE FORMATION (HTCOAG.BAS)

The third computer program, HTCOAG.BAS, models the coagulation of the ultrafine nuclei predicted by NUCL.FOR. The size range from the nuclei size to the final size is discretized into equal volume size intervals. The classical coagulation equations are then used to predict the coagulation growth into and out of each size interval, beginning with the first interval above the nuclei size. The rate equations for all intervals are integrated, and the particle number concentrations in all intervals are updated at small time increments. This is repeated until the particle number concentrations become so low that coagulation is no longer significant or until a specified time period is exceeded. In all of the cases examined to date, the coagulation rate became negligible within a time period of 3 sec after nucleation.

The program HTCOAG.BAS was originally developed by Dolan (30) to model the coagulation of diesel exhaust particles. Dolan reported that the size distributions predicted by this program agreed reasonably well with measurements obtained from the exhaust of a single-cylinder diesel engine. Only minor modifications of the program were necessary to make it applicable to ultrafine fly ash. The mathematical derivation of this model and a listing of the computer code may be found in reference (30). The application of the model to ultrafine fly ash is discussed in more detail in reference (29).

INPUT DATA REQUIRED TO RUN THE PROGRAMS

The input data required to run each of the three programs are listed in Table 1. Most of these parameters are determined through standard coal and ash analyses that are performed in the early stages of power plant design. The more esoteric parameters such as number of char fragments per coal

particle and surface tension of the molten ash can be obtained from the literature. Of course, the first three input parameters listed for HTCOAG.BAS must be obtained from the output of NUCL.FOR. If the results of the calculations are used in the design of a control device, the time interval chosen for coagulation is not critical, since in all cases simulated coagulation growth was found to be complete in less than 3 sec. The residence time between the burner zone and the control device is typically at least 5 sec.

PARAMETRIC STUDIES USING THE MODELS

The three models were used in combination to study the effect of variations in the input parameters on the predicted particle size spectra. This was done by first establishing a base case that was characterized by the input data listed in Table 2. These input data were selected to represent the middle of the expected range of values for each input parameter. Each input parameter was then varied over the range of values that would be encountered in commercial practice, and the effect on the predicted particle size distribution was examined. Only a single parameter at a time was varied, with all other parameters held constant at the base case values. The results of this study are summarized in Table 3 and Figures 3-5. As indicated in Table 3, all of the input parameters were varied, but only the effects of coal ash content, mean coal particle size, and flame temperature were plotted.

Figure 3 shows that increasing coal ash content increases the masses and shifts the modal diameters of both particle size modes towards larger sizes. The growth in the modes is to be expected since more ash is available for both volatilization and coalescence. The shift in the small particle mode is the result of increased nuclei concentrations and increased coagulation growth rates. The coal ash content can be manipulated by changing coal supplies or altering the coal preparation process. The choices of coal supply and coal preparation process are made early in the design of a power plant, so early identification of troublesome coals is essential. A coal might be considered to be troublesome if the submicron particulate mass can be expected to exceed a specified level. This level could be selected based on the ability of the stack gas to meet opacity regulations with an affordably sized precipitator.

The coal particle size affects only the large particle mode as shown in Figure 4. The small particle mode was unaffected because the burnout time supplied as input to the model was held constant (in reality, however, the required burnout time would increase with increasing coal particle size). Coarser grinding of the coal from a mean size of 30 to 70 μm causes a shift in the modal diameter from 8.2 to 19 μm . This could make the ash easier to collect in a marginally sized precipitator (e.g., a precipitator that was sized for an ash resistivity that was lower than that actually encountered). The effect of this change on an adequately sized precipitator that is in good operating condition should be minimal, however.

Figure 5 shows that the flame temperature affects only the small particle mode since the primary influence of flame temperature is on the volatilization rates. The flame temperature could also have some effect on the large particle mode to the extent that it affects the degree of fragmentation of the

char. There is no information on this effect in the literature. For a more complete discussion of the effect of other parameters, the reader is referred to reference (29).

VALIDATION OF THE MODELS

The computer models were validated against data from a full-scale pulverized coal-fired utility boiler and a pilot-scale pulverized coal-fired combustor. The full-scale unit was a 575-MW, tangentially fired, dry bottom, water-wall furnace that was firing a low-sulfur, high-ash, Texas lignite (31). The pilot-scale combustor was a 1,000,000 Btu/hr, down-fired, refractory-lined unit equipped with simulated superheater and reheater sections (32). The coal burned in the pilot-scale unit was a medium-sulfur, low-ash, Indiana bituminous coal. The particle size data for the full-scale unit were obtained as part of a recent evaluation of fabric filter performance (31). The particle size data for the pilot-scale unit were obtained under an EPA-sponsored study of furnace sorbent injection for SO_2 control (32).

FULL-SCALE UTILITY BOILER DATA

Figure 6 shows a comparison of the particle size distributions predicted by the models and measured at the full-scale utility boiler. The circles, triangles, and squares indicate particle size data obtained with an electrical aerosol size analyzer, an optical counter, and a five-stage cyclone assembly. The ash content, ash density, and silica and alumina contents represent the average values obtained from standard analyses reported in the field test summary (31). However, the values of the coal density, number of char fragments, mean coal particle size, Rosin-Rammler exponent, burnout time, and surface tension of molten ash were set equal to the base case values because these data were not obtained during the field test. Even without any adjustment of these parameters, reasonably good agreement between the predicted and measured distributions was obtained. It should be noted, however, that the sharp ultrafine particle mode will appear to be much broader unless the raw data obtained with the electrical aerosol size analyzer (EASA) are corrected for channel cross-sensitivities (9,29). The full-scale utility boiler data were corrected for this effect, but the pilot-scale data, discussed below, were not corrected because the raw data were not available in this case.

PILOT-SCALE COMBUSTOR DATA

Figures 7 and 8 show comparisons of the pilot-scale combustor data to predictions made using two different assumed mass loadings. This was done because about 60% of the total particulate mass had dropped out of the pilot combustor system ahead of the sampling point. Therefore, the normalized mass distribution was converted to a non-normalized form by multiplying by both total mass loadings. The lower loading, 1993 mg/dncm, is the measured value after dropout; the higher loading, 5020 mg/dncm, is the predicted value based on an empirical correlation between mass loading and ash content (33). The dropout should affect only the larger particle sizes since only the large particle fraction does not reach the sampling point. Therefore, it may be

appropriate to apply the lower loading above a few microns and the higher loading below a few microns in particle size. The lower loading does appear to give better agreement in the large particle mode. The fact that the higher loading does not yield better agreement in the small particle mode is probably attributable to the fact that the pilot-scale combustor EASA data were not corrected for channel cross-sensitivities. In any case, the agreement between the predictions and measurements appears to be quite satisfactory.

CONCLUSIONS

The large particle mode is formed by coalescence of molten mineral inclusions that adhere to the receding surfaces of char fragments during burnout. Based on this analysis, it is concluded that the formation of the large particle mode of fly ash is governed by

1. The ash content of the coal
2. The particle size distribution of the pulverized coal
3. The degree of fragmentation of the coal char during burnout
4. The coal and ash densities

The small particle mode of fly ash is formed by volatilization of certain mineral components followed by homogeneous nucleation and coagulation of the nuclei. Bubble bursting does not appear to be a major source of small ash particles because the sharp ultrafine mode is more accurately predicted by the volatilization/nucleation model. The critical factors that influence the small particle mode are

1. The ash content of the coal
2. The flame temperature
3. The composition of the coal mineral matter
4. The coal particle burnout time

Low boiling metals such as arsenic, selenium, mercury, and cadmium are enriched in fine particle size fractions, but they are not the major contributors to the small particle mass. The small particle mode is similar to the large particle mode in terms of their major chemical components.

The difficulty in measuring some of the required input parameters (e.g., char fragments per coal particle or surface tension of molten ash) is not a deterrent to the use of the models since accurate predictions can be made with values taken from the literature. The recommended values are

1. Char fragments per coal particle = 4 (Reference 7)
2. Mean coal particle size = 53 μm (References 25 and 26)

3. Rosin-Rammler exponent = 1.2 (References 25 and 26)
4. Surface tension of molten ash = 320 dynes/cm (Reference 17)

Of course, the actual size distribution of the pulverized coal should be used if available.

FUTURE WORK

It is hoped that the models described here will be incorporated into a precipitator sizing procedure. Such a procedure would be useful in designing precipitators for new coal-fired power plants and identifying potentially troublesome coals. It could also be used to predict the effect of switching coal supplies or changing boiler operating conditions on the performance of an existing precipitator.

The volatilization portion of NUCL.FOR may be adapted to a study of ash deposition on boiler tubes. It is known that such deposits are initiated through the formation of a sticky alkali-rich tube coating that is deposited from the vapor phase (1,2). An understanding of the initiation mechanism depends on an accurate treatment of the alkali volatilization. The latter may be described using an approach analogous to that discussed here.

DISCLOSURE

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

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TABLE 1. INPUT DATA REQUIRED TO RUN THE COMPUTER PROGRAMS

BREAK.FOR

Ash content of coal, %
 Coal density, g/cc
 Ash density, g/cc
 No. of fragments per coal particle
 Mean coal particle size, μm
 Rosin-Rammler exponent
 Total mass loading, mg/dncm

NUCL.FOR

Flame temperature, °K
 Ash content of coal, %
 Silica content of ash, %
 Alumina content of ash, %
 Burnout time, sec
 Surface tension of molten ash, dynes/cm²

HTCOAG.BAS

Number concentration of ultrafine nuclei (predicted for NUCL.FOR), cm⁻³
 Mean size of ultrafine nuclei (predicted by NUCL.FOR, μm
 Standard deviation on mean size
 Ash density, g/cc
 Time interval for coagulation, sec

TABLE 2. INPUT AND OUTPUT DATA FOR BASE CASE

Input Data for Base Case

Coal ash content, %	15
Coal density, g/cm ³	1.3
Ash density, g/cm ³	2.4
Char fragments per coal particle	4
Mean coal particle size, μm	50
Rosin-Rammler exponent	1.2
Total mass loading of fly ash, mg/dnm ³	10,307
Flame temperature, °F/°K	2,700/1,756
Silica content of ash, %	50
Alumina content of ash, %	20
Coal particle burnout time, sec	0.6
Surface tension of molten ash particles, dynes/cm	320

Output Data for Base Case

Ultrafine mode, peak dM/dlogD value (mg/dnm ³)	2.6×10^{-1}
Ultrafine mode, modal diameter (μm)	0.012
Supermicron mode, peak dM/dlogD value (mg/dnm ³)	1.1×10^4
Supermicron mode, modal diameter (μm)	13.6

TABLE 3. SUMMARY OF MODEL SENSITIVITY TO INPUT PARAMETERS

Input parameter	Range of variability	Mode affected*	Range of peak dM/dlogD values+	Range of modal diameters**
Coal ash content, %	5-30	supermicron	3.7×10^3 to 2.1×10^4	10-18
		ultrafine	0 to 1.5×10^3	0-0.15
Coal density, g/cm ³	1.1-1.5	supermicron	1.1×10^4	12.9-14.3
Ash density, g/cm ³	2.0-3.0	supermicron	1.1×10^4	14.5-12.7
Char fragments per coal particle	3-5	supermicron	1.1×10^4	15.0-12.7
Mean coal particle size, μm	30-70	supermicron	1.1×10^4	8.2-19
Rosin-Rammler exponent	1.0-1.4	supermicron	8.7×10^3 to 1.2×10^4	13.6
Flame temperature, °F °K	2300-3100 1533-1978	ultrafine	5.9×10^{-2} to 1.6×10^0	0.0075-0.021
Silica content of ash, %	30-70	ultrafine	0 to 3.5×10^{-1}	0-0.013
Alumina content of ash, %	10-30	ultrafine	0 to 2.2×10^0	0-0.017
Burnout time, sec	0.4-0.8	ultrafine	3.0×10^{-4} to 5.6×10^0	0.0073-0.022
Surface tension, dynes/cm	200-400	ultrafine	6.7×10^{-1} to 5.6×10^{-2}	0.014-0.0071

*Only the mode or modes that are affected by the indicated input parameter are listed here.

+Peak dM/dlogD values given in milligrams per dry normal cubic meter (mg/dnm³).

**Modal diameters given in microns.

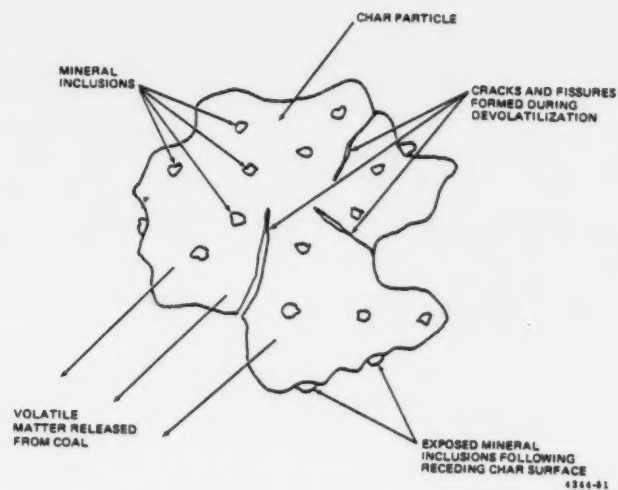


Figure 1. Conceptual drawing of residual char particle after devolatilization of organic matter from coal.

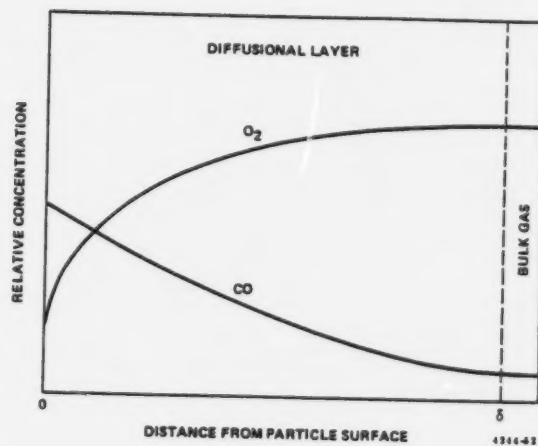


Figure 2. Hypothetical concentration gradients near char particle surface.

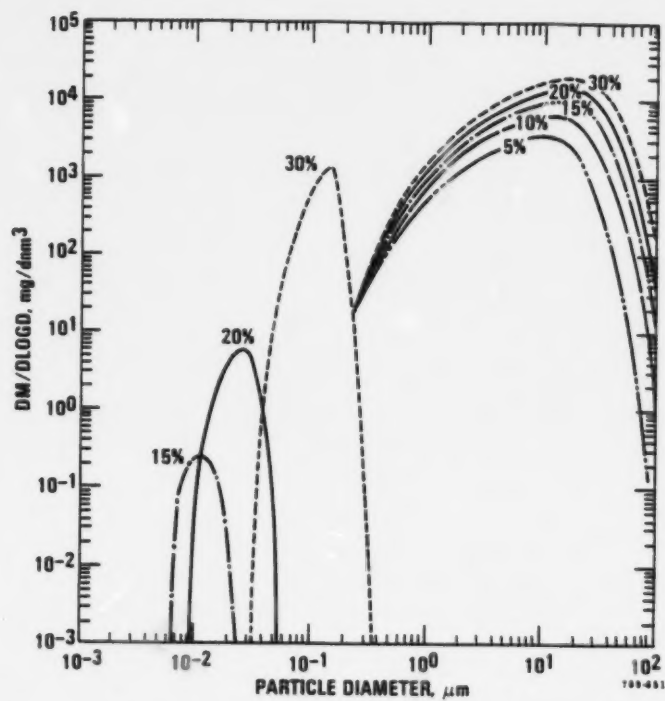


Figure 3. Effect of ash content on the ultrafine and supermicron modes of the fly ash particle size distribution.

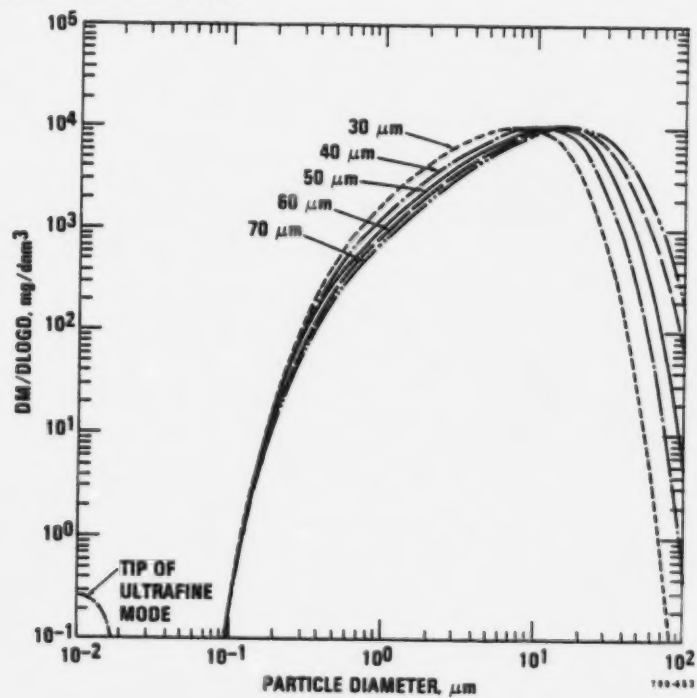


Figure 4. Effect of mean coal particle size on the supermicron mode of the fly ash particle size distribution.

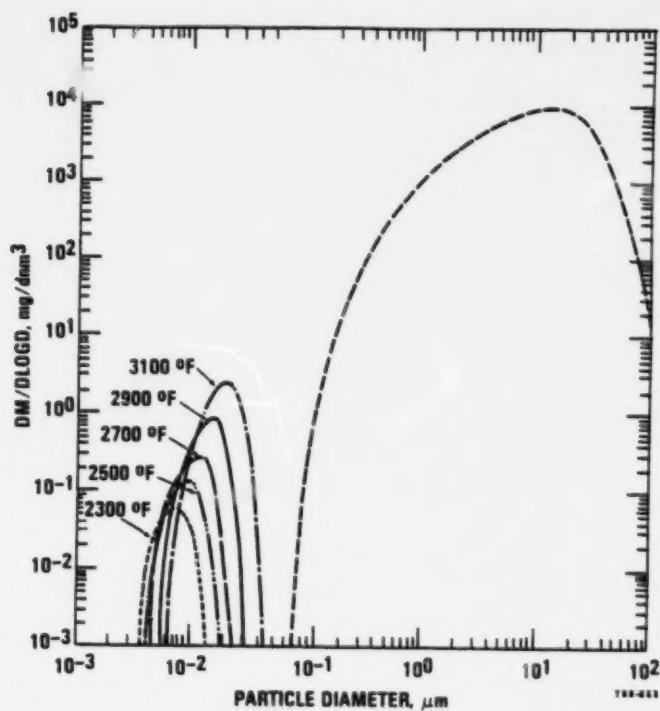


Figure 5. Effect of flame temperature on the ultrafine mode of the fly ash particle size distribution.

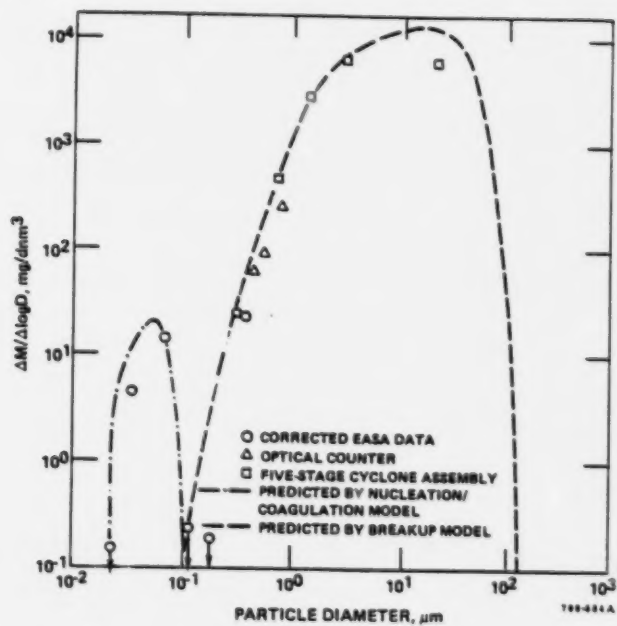


Figure 6. Comparison of model prediction and size distribution measurements at a full-scale pulverized coal-fired power plant, after correction for cross-sensitivities in electrical aerosol size analyzer.

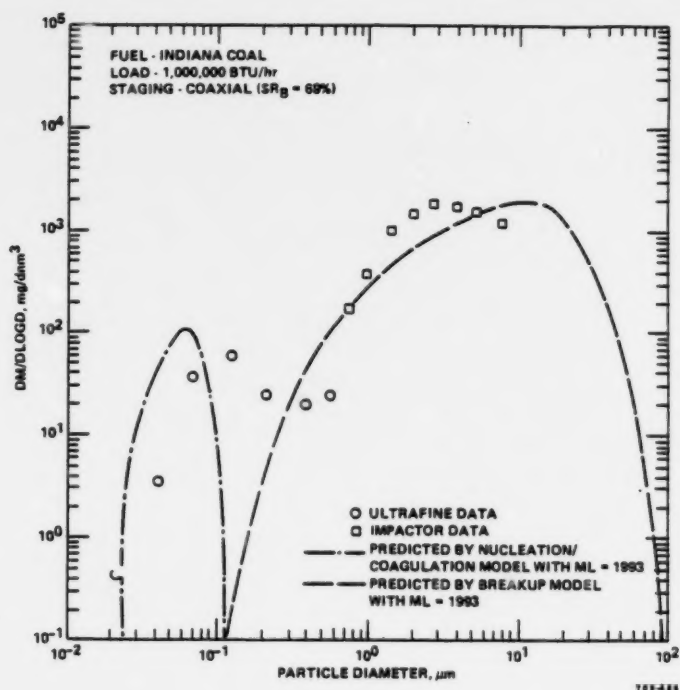


Figure 7. Comparison of model prediction, assuming lower total mass loading, with size distribution measurements at pilot combustor.

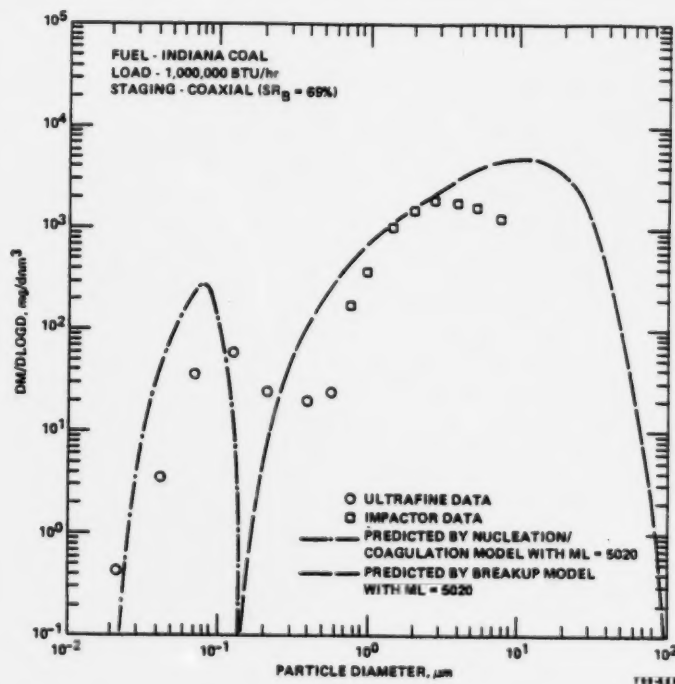


Figure 8. Same comparison shown in Figure 7 except with higher total mass loading.

QUALITY ASSURANCE CONSIDERATIONS FOR PARTICLE SIZE MEASUREMENTS

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ABSTRACT

All measurement activities conducted by or in behalf of the Environmental Protection Agency are required to implement quality control and quality assurance techniques to ensure that the quality of the results of the measurements may be determined and their adequacy assessed and documented. Traditional indicators of data quality include precision, accuracy (or bias), completeness, representativeness, and reproducibility. For routine measurement systems, such data quality indicators are readily achievable; however, particle size measurements pose special limitations, which suggest that such measurements are non-routine and may require non-traditional approaches to determining the quality of the data. The lack of a Reference Method and Certified Reference Materials may make precision and bias, respectively, particularly difficult to determine.

This paper will discuss approaches to determine adequate data quality indicators for particle size measurements and the limitations or uncertainties with such measurements. The focus will be on in-situ particle measurements using cascade impactors and cyclones. Its intent is to provide an increased awareness of the quality problems associated with these measurements and to provide guidance on considerations for test plan design, QA Project Plan preparation, and self-auditing techniques for users of impactors and cyclones.

This paper has been reviewed in accordance with the U.S. Environmental Protection Agency's peer and administrative review policies and approved for presentation and publication.

BACKGROUND

The data quality requirements for environmentally related measurements obtained under the auspices of EPA range from qualitative estimates from exploratory research to the rigorously documented quantitative information to support regulation development or litigation. These apparently disparate requirements have one thing in common: all these data are collected to produce information for a specified purpose. Once this purpose has been defined and appropriately communicated, a quality assurance (QA) system may be developed and implemented to ensure that the information meets the required data quality specifications.

Traditional indications of data quality include precision, accuracy (or bias), representativeness, completeness, and comparability, which are often abbreviated PARCC. Consequently, data quality objectives have been expressed in terms of PARCC. It is EPA policy that all measurement related activities will have their data quality objectives documented in a Quality Assurance Project Plan. That is, a written plan must be developed that will set forth the objectives or criteria by which the quality of the technical results will be judged. In this way, the QA Project Plan may be utilized to develop a test program with the necessary quality checks to ensure that the data generated are defensible and are adequate for the original intended use of the data.

The intended use of the data is the key criterion from which all data quality criteria or objectives are derived. For most routine measurements, the state-of-the-art methodology is such that data quality indicators (e.g., PARCC) may be determined readily. For particle size measurements, the applicability of PARCC as data quality indicators is not so clear. The susceptibility of these measurements to external perturbations may make the specification and determination of PARCC values difficult. It is important, therefore, to examine the entire particle size measurement process to identify the proper application of conventional data quality indicators.

The remainder of this paper will focus on problems associated with the determination of data quality indicators for particle size measurements. The intent of this discussion is to increase the awareness of these problems by identifying areas in which data quality issues should be of concern. In particular, guidance is provided for consideration in test plan design, QA Project Plan preparation, and data analysis techniques available to users of impactors and cyclones, but with particular emphasis on impactors.

TEST DESIGN

The initial phase of a successful, cost-effective measurement program is the definition of the technical question(s) to be answered. The more clearly these questions can be stated, the higher the probability that an experiment can be designed that, when implemented, will result in clear and defensible answers. The definition phase is usually a difficult one, yet the struggle to define the question(s) concisely will in itself assist in the definition

process. The reward is that the results of the measurement process (i.e., data) will be of sufficient quality to ensure that the answers to the technical questions posed will be adequate for their original intended use. That is, the quality of the up-front planning will largely determine the quality of the technical results achieved and, ultimately, their utility to the user.

DATA QUALITY OBJECTIVES

Typical measurement programs focus on the issues pertaining directly to sampling and analysis and not on the design of the test or experiment itself. Specific issues relating to the test design phase are given in Table 1. The planning phase is where the objectives for the data quality indicators must be explicitly addressed. That is, how are sample collection and sample analysis methods to be selected and implemented so that accuracy is maintained and documented within acceptable limits? How is the sampling strategy developed to ensure it is representative of the intended population? What level of completeness will ensure not only that the data will support the required confidence limits, but that non-uniformity in the sample rejection does not unduly cause unrepresentativeness of important subpopulations. These considerations should be clearly and concisely described in a manner that allows the final results to be readily compared against them.

Finally, it is essential to consider the potential audience to whom the final information or report will be presented. The end user must be able to understand the information provided. The key considerations for presentation of information are given in Table 2.

The goal of the planning exercise is to anticipate communication of the answers to "real questions" based on the measurement efforts. This communication must include evaluation of the estimated reliability of the answers as an integral component; that is, are the results credible?

Following preliminary planning, the development of a test plan becomes more quantitative. In the case of particle size measurements, each aspect of the test procedure should be assessed to determine its potential impact on data quality. Unfortunately, the impacts are not always obvious. Moreover, several factors may work together to create adverse data quality impacts. For example, if a high temperature stack gas is to be sampled, the user must have a probe, such as a quartz probe, capable of withstanding high temperature. In addition, the gas may be highly reactive, so other parts of the sampling train may require the use of non-reactive materials. Users should not be lulled into a false sense of security regarding particle sampling. It is well known that impactors and cyclones can be added to the front end of a Method 5 sampling probe. The operation of such a system according to the Method 5 protocol requires isokinetic conditions, which usually requires the operator to vary the gas flowrate through the sampling train. While this is entirely appropriate for Method 5, it has a catastrophic effect on the impactor measurements that require a constant flowrate through the device in order to calculate the size cut point of the impactor.

Five principal areas of particle size sampling must be considered in terms of their impacts on the overall quality of the data produced. These are: source characteristics, equipment limitations, personnel training and skill, operational logistics, and data reduction and analysis. Considerations associated with each of these activities are summarized in Table 3.

QUALITY ASSURANCE PROJECT PLAN

Guidelines for particle size test preparation have been documented in the literature⁽¹⁾. These include equipment selection, calibration, operation, and sample data reduction and analysis⁽²⁾. They are related to data quality through the Quality Assurance Project Plan (QAPP), which is required to be approved prior to the start of measurement activities in EPA. The principal elements of an EPA QAPP are given in Table 4. As can be seen, the QAPP addresses the major parts of particle size source tests. The content of the QAPP will depend largely upon the complexity of the test program and the intended use of the data. The QAPP should be regarded as a companion document to the formal test plan that integrates the test design with consideration of required data quality.

The preparation of the QAPP depends on an adequate test design. If the test plan is quantitative, then the QAPP should be quantitative as well. As noted earlier, the problems of quantifying traditional data quality indicators for particle size measurements make this task difficult. This will be seen particularly in regard to equipment calibration, which will be discussed later.

Careful selection and preparation of the measurement equipment strongly affects the two quantitative aspects of the data quality indicators--accuracy (or bias) and precision. Selection of the impactor (or cyclone) design and understanding its unique limitations are essential for impactor measurements simply because reference materials are unavailable for use in a dynamic, realistic test system. Laboratory calibration, using test aerosols of a known size, is a positive step, but it is far different from the operating environment of the working impactor, especially in terms of temperature and particle size distribution. Wall effects, substrate capture efficiency or reintrainment, and gas stream interactions with substrate materials all contribute to bias the calculated distribution. Each may be detected or avoided by careful experiment design. Regular use of blank impactor runs (that is, impactors that are identical to those used in the stack sampling but are not exposed to the stack gases) frequently used to identify and quantify bias and imprecision. These blank runs require resources that are generally much less than the cost of indefensible or poor quality data.

Ancillary equipment, while not a prominent consideration in the planning phase, may contribute quite significantly to undetectable experimental bias if not explicitly checked. For example, minor defects in the nozzle tip may alter the inlet area of the nozzle resulting in a significant departure from isokinetic conditions in the experiment. The apparent or measured size distribution is thus altered in an unquantified manner, which introduces undetected systematic error (bias) into the final results. Similarly, maintenance and calibration of the pump and dry gas meter play crucial roles in causing undetected bias in terms of leakage and inaccurate volume measurements. The probe, probe

liner, and umbilical line must be evaluated for their effect on experimental bias. Such issues as interactions with the gas, temperature sensitivity, and flow restriction must be considered not only as part of the experiment, but as part of the quality control examination of the test plan design documented in the QA Project Plan.

The importance of carefully evaluating each component of the sampling system can be seen when particle size is to be measured. Documentation of these considerations in the QA Project Plan provides a "road map" to achieving the needed quality of the results. The actual implementation of quality control (QC) may be seen best by examining the elements of the test itself with respect to how data quality is impacted by each element and what can be done to ensure adequate quality. These elements include pretest preparation, sample collection, sample handling, sample analysis, data reduction and analysis, and information presentation.

PRETEST PREPARATION

The quality control aspects of the QA Project Plan must focus on the detailed sequence of events in the test activity that contribute to the overall quality of information produced by the test. Several activities critical to the overall data quality must be performed at the test site or prior to the test site to ensure that the data quality objectives are achievable.

EQUIPMENT CALIBRATION

Among the key pieces of equipment to be reviewed and calibrated are the dry gas meter and the impactor itself. The calibration standards, and the conditions under which they are used in the calibration procedure, must be rechecked in the same manner. For example, the distribution of jet diameters for the stages should be checked against earlier values. Changes from the values used in the data reduction procedure will affect the ultimate interpretation of the results. Calibration of an impactor under simulated test conditions may be costly and may not yield defensible results in every situation. Yet the specification of the data quality objectives may indicate that such is necessary. Dry gas meter calibration is a routine operation, since it is part of the EPA Method 5 procedure. Other, less obvious, calibration requirements may include nozzle quality checks and sampling train leak checks similar to those of EPA Method 5. These should be discussed in the QA Project Plan.

SOURCE OPERATING CHARACTERISTICS

Source operating characteristics also affect data quality. In some cases, the source environment to which the impactor operators are exposed may impact negatively on data quality. Harsh test conditions may require more-than-minimal operator shelter if the quality of sampling train operations is to be maintained adequately. Fluctuations in the duct gas velocity and, to a much lesser extent, gas composition have direct, but difficult to quantify, implications for the quality of the reduced data. However, such fluctuations may be addressed during data reduction if criteria for determining the error associated

with the fluctuations are identified in the QA Project Plan, and are subsequently implemented during data reduction.

SAMPLE COLLECTION

The first interaction of the sampling system with the stack gas environment is at the impactor probe nozzle. The probe typically enters a duct that is not at ambient pressure or temperature, in which a possibly noxious gas is flowing, and contained by walls that are coated with heterogeneous waste particles of unknown origin and description. The velocity profile across this duct is rarely "flat." It may be skewed due to nearby turns in the duct or to changes in the duct size, which may induce on-site alteration of the test plan sampling strategy. This alteration must simultaneously address the conflicting data quality requirement for representativeness (i.e., sampling across the duct) and the technical requirement for constant impactor flowrate at isokinetic conditions. The magnitude of this challenge is routinely identified using standard Method 1 procedures, yet ensuring representativeness is difficult with currently available impactors. And, in the process, the nozzle must be positioned accurately at precalculated points along the duct diameter while ensuring that probe and sampling port interactions do not disturb the duct flow pattern and gas composition significantly.

SAMPLE HANDLING

Once sampling has been completed successfully, the processing of the impactor sample begins with its handling and transfer through an appropriate sample conditioning process (e.g., desiccation) to weighing, and possible physical and/or chemical analysis. Mechanical disturbances of the tiny deposits of particulate matter must be anticipated and prevented, either in impactor disassembly or in storage and transport. Electrostatic effects are inherent in the collection process, especially in low moisture gas streams, causing the particles to actually jump off the substrates in severe situations. Brushdown of the impactor, if improperly performed, may contribute to undetected wall losses of particulate matter, which will result in the introduction of undetected bias in the test data. Brushdown of the nozzle and probe (ahead of the impactor itself) is responsible for capturing many large particles, which contribute a disproportionately large fraction of the total sample mass. Improper substrate conditioning may alter the sample or bias the apparent sample weight by the unaccounted for adsorption or desorption of moisture or low molecular weight substances from the exposed substrate. Again, appropriately used blanks must be utilized to detect such problems.

SAMPLE ANALYSIS

Impactor sample analysis requires weighing the substrates under controlled conditions, both before and after they are exposed to the gas stream. The substrates typically weigh 100 milligrams (mg), and the typically quoted change is 0.01 mg. The implicit precision is, thus, on the order of 0.01 percent, which is more characteristic of a controlled research laboratory environment than the stack of a stationary source. For this reason, operator training, adequately balanced environment, and carefully reasoned standard

operating procedures (SOPs) offer the best hope for obtaining weight data of reasonable quality.

In addressing the issue of sample data quality, and to minimize overall test costs, weighing is sometimes done at the home laboratory. In these situations, additional considerations of the effect of handling, transportation, and other variables on the substrates and particles must be addressed. For example, volatilization of unrecognized low molecular weight particles would lead to a time-and/or size-dependent weight bias. If a suitable environment can be created at (or very near) the sampling site, many of these effects can be minimized to an acceptable level. An added advantage of this option is that data can be reduced on-site for tests of 3 days or longer. Blank impactor runs are again a necessity in estimating the magnitude of this bias.

DATA REDUCTION AND ANALYSIS

Collection of data from which data quality may be estimated requires foresight and determination. For example, it may seem difficult to justify blank runs, comparable in all operating conditions to data runs, of at least daily frequency. Similarly, replicate runs, at the "same" source operating conditions, may be regarded as unnecessary duplication. Yet, it is precisely such runs that permit the estimation of the total system precision of the final results, and that permit the evaluation of the data against the original data quality objectives. Similarly, checks of operating conditions (e.g., pressure, temperature) and laboratory parameters (e.g., weights) for their quality are necessary to permit estimation of the quality of the results.

Stage cut points typically are calculated using the D50 approach, where it is assumed that 50 percent of the fraction of particles with the specified diameter are captured. The other 50 percent pass to the next stage where they, too, are then captured. It is widely recognized that this theory does not adequately describe the operation of real impactors, which do not exhibit such simple behavior. This assumption does, however, permit routine, comparable calculations under working conditions. Even under this simplifying assumption, non-uniformity of the parameters from which the cut points are calculated (e.g., temperature, pressure, jet diameter, flowrate) contributes to inaccuracy and imprecision in these intermediate results. For example, flow through the jets cannot be simply calculated as average flow through the impactor due to distribution of jet diameters on a given stage, plugging of jets, etc. Impactors so configured capture significant particulate matter on stages that are preceded in the gas flow by a stage of identical cut point. This overlap of catches on identical stages must be addressed in the data reduction process, and is especially dependent on the design of the impactor being used. Operating conditions not only must be monitored during sampling, but must be reviewed during data analysis in order to assess the possibility of induced sampling errors, such as reentrainment or blowoff of captured particles from stage to stage.

Criteria for identifying and selecting "equivalent" impactor runs should be clearly established in the QA Project Plan to support the data quality objectives which have been set. Once individual impactor runs have been reduced to stage cut points and loadings, the issue of the comparability of

these individual runs to the overall particle size distribution presentation must be addressed. Criteria for detecting, evaluating, and treating the effect of "outlier" runs must be established separately from the treatment of runs where known operating errors occurred. In regard to reproducibility, it is impossible to ever repeat an impactor experiment on a real source with exactly the same conditions in the stack. Thus, the traditional concept of reproducibility may not be achievable for particle size measurements.

Completeness may also be affected if conditions during the test activity prevented the acquisition of all of the data or samples necessary for the technical decision to be made. The data quality objective for completeness will define the minimum amount of data needed. If the completeness objective was not attained, then the data analysis phase must address the implications of such non-attainment on the overall data quality.

INFORMATION PRESENTATION

The final step in the process of developing information from data is the presentation format. The concern here is a classical problem of communication: to present information so that the user can accurately comprehend the summarized results of the test. This requires explicit consideration of the audience and the intended use of the data. In particle size measurements, the complexity of the information is such that the format may need to vary, depending on the audience. A scientist may need details that are not needed by another user of the data. In addition, the intended use of the data will impact its presentation.

Methods for presenting technical information rapidly and inexpensively are slowly developing. Traditional tables and graphs in bound reports are giving way to computer-generated artwork and tables that increase the flexibility of the user to offer the information to the reader in a variety of styles. Reduced data with different interpretations can be easily, quickly, and inexpensively depicted quite accurately in an easily comprehensible form. Such considerations are the final link in the process of generating and communicating data of known and defensible quality to the end-user. Great care must be exercised in this phase of the work. It will do no good to have done the best possible QA and QC job on the project, only to have the information presented to the user in a form that is confusing or that fails to answer his/her question. The true measure of adequate data quality lies in the utility of the data to the end-user.

SUMMARY

Data quality considerations impact virtually every aspect of particle size measurements. In fact, few other environmental measurements are as susceptible to adverse quality impacts from outside perturbations as particle sizing. The adequate implementation of QA and QC requires a systematic assessment of the test data to ensure that the data quality needed to answer the technical question posed is, in fact, achievable and documentable.

The application of traditional data quality indicators may not be possible; however, alternatives should be specified in a QA Project Plan even if the objectives are not quantitative themselves. The QA Project Plan will

provide a means of assessing the effectiveness of the applied QC program and a measure of the attainment of quality goals.

Finally, it is not enough to do a "good" job of gathering the test data and calculating tables of numbers. The information produced must be presented to the user in a manner that he/she can understand and use. The final measure of how "good" the measurement data are will be determined by the usefulness of the information provided.

REFERENCES

1. Harris, D. B., "Procedures for Cascade Impactor Calibration and Operation in Process Streams," EPA-600/2-77-004, U.S. Environmental Protection Agency, Research Triangle Park, NC, January 1977.
2. Tatsch, C. E., W. M. Yeager, and G. L. Johnson, "PADRE: A Computerized Data Reduction System for Cascade Impactor Measurements," Journal of the Air Pollution Control Association, 34(6):655, June 1984.

TABLE 1. DATA QUALITY RELATED ISSUES DURING TEST DESIGN

Issue	Relation to Data Quality
WHAT	What is the problem to be solved? What is the intended "population" to which the results will be applied?
WHY	Why are the data needed?
HOW	How "good" do the results have to be?
WHO	Who will use the results?

TABLE 2. DATA QUALITY CONSIDERATIONS FOR INFORMATION PRESENTATION

Issue	Relation to Data Quality
FORMAT	What is the most readily understandable format for communicating the results?
CLARITY	Does the user really understand what the technical results say?
RELEVANCE	What parts of the information provided contribute to answering the questions the user thinks are important?

TABLE 3. SAMPLING AREA IMPACT CONSIDERATIONS FOR DATA QUALITY

Activity Area	Considerations for Data Quality Impact
SOURCE CHARACTERISTICS	Are there any unusual source operating conditions? How is the source being operated (e.g., steady state, cyclic, intermittent)? What is the environment inside the stack (e.g., temperature, flowrate)?
EQUIPMENT LIMITATIONS	What type of sampling equipment is available (e.g., impactor, cyclone)? Are the operating ranges of meters, instruments, etc., adequate for the source conditions?
PERSONNEL TRAINING/SKILL	Does the sampling crew have adequate experience for this particular source test? Are any special physical requirements needed for this job?
OPERATIONAL LOGISTICS	Is the sampling location accessible? What sample handling arrangements must be made?
DATA REDUCTION AND ANALYSIS	What computations are required to complete the analysis of the data? What are the criteria for the rejection of outlier data? How are the computations checked for accuracy?

TABLE 4. ELEMENTS OF A QUALITY ASSURANCE PROJECT PLAN

Element	Description
1	Project Description (including intended use of the data)
2	Project Organization and Project QA Organization
3	Data Quality Objectives for Critical Measurements
4	Site Selection and Sampling Procedures
5	Sample Custody
6	Calibration Procedures and Frequency
7	Analytical Procedures
8	Data Reduction, Validation, and Reporting (including how missing or outlier data are to be handled)
9	Internal Quality Control Checks and Audits
10	Plans for Performance and System Audits
11	Preventive Maintenance (procedures and schedules)
12	Calculation of Data Quality Indicators
13	Corrective Action (criteria and procedures)
14	QC Reports to Management

Session C-3: ADVANCED ENERGY APPLICATIONS I

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TECHNICAL AND ECONOMIC EVALUATION OF HIGH TEMPERATURE
AND HIGH PRESSURE PARTICULATE CLEANUP SYSTEMS

by

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ABSTRACT

Gilbert/Commonwealth has been under contract to DOE/METC for the past five years providing technical assistance for developing high temperature, high pressure (HTHP) gas cleanup devices. Responsibilities have included test monitoring, technical and economic evaluations and test facility design.

The status of HTHP particulate cleanup will be presented, including DOE, EPRI and efforts overseas. Emphasis will be on the DOE program where a significant development effort is being conducted.

Results of a recent technical and economic evaluation of ten commercial sized HTHP systems for pressurized fluidized bed technology will be presented. Ten devices have been tested by DOE from bench to rig scale levels at various conditions. The past technical achievements and the future test program are discussed as well as the potential for scaleup.

EPRI's involvement with woven and hard ceramic filters will be reviewed, including plans for the largest scale HTHP test to date. A timetable for potential commercialization of HTHP concepts as well as an assessment of this potential will be presented.

INTRODUCTION

The U.S. Department of Energy (DOE) has funded the research and development of several devices designed to remove particulates from high temperature, high pressure (HTHP) Pressurized Fluidized Bed Combustion (PFBC) exhaust gases. With many test programs completed, and others underway, a comprehensive evaluation, with commercial perspective, was undertaken by Gilbert/Commonwealth under contract to the DOE, Morgantown Energy Technology Center (METC), located in Morgantown, WV (Contract No. DE-AM21-82MC19196).

DOE/METC is continuing the development of certain devices at the rig scale and the choices made were partly a result of the G/C evaluation.

EPRI concurrently has funded the development of HTHP cleanup devices by conducting ceramic bag and porous ceramic filter tests under simulated PFBC conditions and plan to test under actual conditions at the largest PFBC facility located in Grimethorpe, England.

The ten devices under development and their respective manufacturers chosen by DOE for the G/C evaluation follow:

- o Electrostatic Precipitator - Cottrell Environmental Sciences (CES)
- o Ceramic Felt Filter - Acurex Corp.
- o Ceramic Cross Flow Filter - Westinghouse
- o Shallow Static Granular Bed Filter - Ducon/Westinghouse
- o Electrostatic Granular Bed Filter - General Electric (GE)
- o Moving Granular Bed Filter - Combustion Power Co. (CPC)
- o Dry Plate Scrubber - Air Pollution Technology (APT)
- o Magnetic Granular Bed Filter - Exxon
- o Electrocyclone - General Electric
- o Acoustic Agglomerator - Aerojet/Penn State

Each device or system has reached a different stage of development. In the commercial HTHP designs which exist, the design approach has varied in the basic assumptions as well as many other aspects of design and cost.

The objective of this analysis was to provide a technical and economic evaluation of the stated ten HTHP systems for the purpose of prioritizing them according to performance, cost and general viability of achieving commercial status. Ten HTHP system designs were required which approximated the approach that would be taken by a single designer applying a similar degree of conservatism. The purpose of this paper is to summarize the approach and present the results with an emphasis on the unique aspects of the evaluation.

The entire study has been published by the DOE as Report No. DOE/MC/19196-1654 and is available through NTIS.⁽¹⁾

The following activities were elements of this evaluation:

- o Review/Normalize test experience.
- o Normalize equipment designs and configurations.
- o Develop a methodology for assessment of risks vs payoffs.
- o Identify R&D needs for each device.
- o Develop normalized capital and operating costs for each system.
- o Perform engineering trade-off studies.
- o Perform an evaluation consisting of inputs from qualified and utility oriented engineers.

The scope included modification to several existing HTHP cleanup system designs, as well as an attempt to make cost minimizing design choices. Each supplier of HTHP cleanup equipment provided input to the normalized design, costs, performance, past and future test programs, and R&D needs for their

device. General agreement was reached regarding the normalization. Experts in HTHP cleanup fundamentals were consulted to assist in developing a method to normalize test data from the HTHP devices.

HTHP PARTICULATE CLEANUP DEVELOPMENT STATUS

During the course of the testing and development of the HTHP devices, different arrangements, models and concepts were evaluated. Full scale designs cannot yet be projected with total confidence and concepts may change significantly as the devices are subjected to more severe test conditions. Tests have been conducted in a myriad of facilities at varying conditions and with mixed results. Further, none of the devices have been operated for even 200 hours at HTHP conditions to demonstrate reliability. Tests of at least 1000 hours, and preferably 5000 hours are needed to demonstrate reliability.

Progress of the devices regarding HTHP operation is summarized in Table 1.(2) The largest size and most extreme operating conditions experienced by the devices are represented. As shown, only five devices were tested on an actual PFBC facility. Nine of the devices were operated at temperatures above 1500° F (815° C); seven experienced high pressure operation; six have achieved NSPS requirements (normalized); and only two have done so in an actual PFBC environment without experiencing significant mechanical failures.

TABLE 1
HTHP TEST SUMMARY

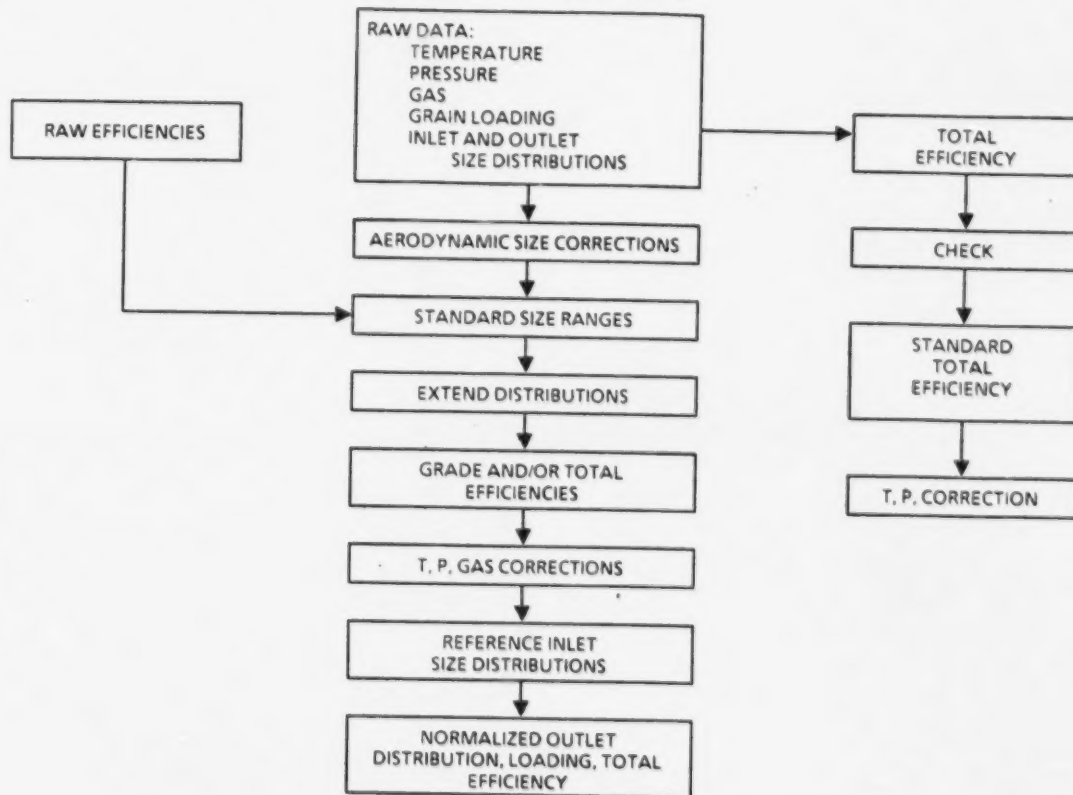
Device	PFBC Type	Maximum Flow, m ³ /min	Velocity, m/min	Maximum Temp, °C	Maximum Pressure, kPa	Collection Efficiency, %	Test Duration, hrs.
Cottrell Environ. Sciences, ESP	Actual	31	28	865	655	72-99.6	362
	Simulated	4.4	53	900	965	83-98.9	>150
Acurex Ceramic Felt Filter	Actual	23	1.5	790	585	99.9	140
	Simulated	14	4.6	815	1140	99.0	77
West. Ceramic Cross Flow Filter	Actual	0.42	3.8	800	930	99.9	90
	Simulated	0.57	5.2	815	1035	99.99	450
Ducon/Westinghouse Static GBF	Simulated	14	14	845	965	77-99	50
General Electric Elect. GBF	Actual	14	15	845	100	70	75
	Simulated	28	12-61	Ambient	100	-	-
Combustion Power Co. Moving GBF	Simulated	46	24	875	100	99.1	200
Air Pollution Tech. Dry Plate Scrubber	Simulated	0.62	19	600	965	70-99.7	0.2
Exxon Magnetic GBF	Actual	1.3	88	880	NA	97.5	-
General Electric Electrocyclone	Simulated	97	2560	885	650	93-96	16
Aerojet/PSU Accoustic Agglom.	Simulated	-	-	Ambient	100	-	-

NORMALIZATION OF TEST DATA

The conditions at which test data have been taken range from ambient to elevated temperature and pressure. In addition, the devices were tested with a wide range of inlet particle loadings and distributions. Therefore, it was desirable to normalize the various test data to reference conditions so that a comparison could be made on an equivalent basis.

The data normalization process proceeded as indicated in Table 2. A computer code was written to implement many of the functions. A method of correcting experimental grade efficiencies based on fundamental collection mechanisms was applied to all devices. Referring to Table 2, raw data is entered; aerodynamic diameters are corrected to physical diameters and these physical inlet and outlet loadings are redistributed in standard size ranges and extended as necessary. Grade efficiencies are determined from the redistributed inlet and outlet size distributions. These, along with reported experimental grade efficiencies, are corrected to reference plant conditions and are applied to appropriate inlet size distributions for the various HGCU devices to obtain an overall device efficiency.

TABLE 2 - DATA FLOW SHEET



The resulting normalized total efficiencies are reported in Table 3. An efficiency of 99.89 percent is required to meet NSPS requirements. The normalized results should be used with caution due to the scarcity of replicate test data and the considerable adjustment to HTHP conditions required for some devices. The best interpretation of Table 3 is to consider the results in groups and not on an individual basis. The test details, the size and configuration of tested hardware, and the potential for improvement must all be considered. As additional and better data is obtained, the normalization process will become a valuable tool.

TABLE 3
HGCU TOTAL PARTICULATE CLEANUP EFFICIENCY (PERCENT)

<u>Device</u>	<u>As Reported by Contractor</u>	<u>Normalized HGCU System with Standard Operating Conditions, Inlet Distributions and Loadings</u>
Ceramic Cross Flow Filter (Westinghouse)(c)	99.96	99.99
Ceramic Felt Filter (Acurex)(c)	99.60	99.98
Moving GBF (CPC)	99.07	99.93
Shallow Static GBF (Ducon/Westinghouse)	98.33	99.96
Electrostatic Precipitator (CES)	98.0	99.6
Magnetic GBF (Exxon)(c)	97.47	99.92
Dry Plate Scrubber (APT)	96.59	99.58
Electrocyclone (GE)	94.3	99.5
Electrostatic GBF (GE)(c)	85.98	98.19
Acoustic Agglomerator (Aerojet/PSU)	81.86(a)	99.96(b)

Notes:

- a) After passing PFBC outlet flyash through a 6 ft. Stairmand Type High Flow cyclone and a 4 ft. Van Tongeren Type High Efficiency cyclone in series.
- b) After passing PFBC outlet flyash through the Acoustic Agglomerator, a 6 ft. Stairmand Type High Flow cyclone and two 4 ft. Van Tongeren Type High Efficiency cyclones in series.
- c) Only data point is single total efficiency; a mean inlet particle size was used.

HGCU SYSTEM DESIGN NORMALIZATION

The purpose of system design normalization is to provide, for each of the HGCU concepts, a scaled-up commercial design which reflects a consistent design approach. Once this design has been finalized, capital costs can be developed, plant performance can be quantified, and trade-off studies can be performed.

The procedure used to develop normalized designs with a consistent design approach was as follows:

- o Establish a reference PFBC commercial plant design.
- o Establish normalization criteria.
- o Collect all pertinent information from vendors, DOE, and literature.
- o Investigate operating parameters, precleanup requirements, materials of construction, auxiliary requirements, heat and pressure losses, and maintenance requirements.
- o Standardize, where possible, precleanup configurations, ash collection/handling, materials of construction, auxiliary systems, pressure vessel designs, and system arrangements.
- o Adjust, where necessary, operating parameters, device configurations, and maintenance schedules.
- o Develop plan and elevation drawings.
- o Determine plant/HGCU system thermal performance.
- o Review the designs with contractors.

The reference plant chosen was a steam-cooled PFBC combined-cycle power plant modified from a GE design. The plant consists of three combustors which discharge gas at 1650° F and 139 psia to the HGCU systems. The steam turbine-generator produces 527 MWe and the gas turbine-generators produce 145 MWe. Net plant output is 662 MWe. Conventional cyclones were used for gas cleanup with a baghouse downstream of the gas turbine to meet NSPS. It should be stressed that the use of cyclones in the reference plant does not imply that cyclones are a "solution" to the HTHP problem but rather are a point of departure. Comparison of the ten HGCU system results to the reference plant should be made considering that a cyclone system may not be suitable, and that an advanced system would be needed to meet both NSPS and turbine protection requirements.

An example of the end result of the design normalization is shown in Figure 1 which depicts a plan and elevation of an electrostatic precipitator used as the HGCU device in the reference plant. A rig scale ESP very similar in design will be tested soon at the NYU PFBC facility.

One of the interesting and original parts of the design normalization was to determine potential maintenance costs of the advanced HGCU system.

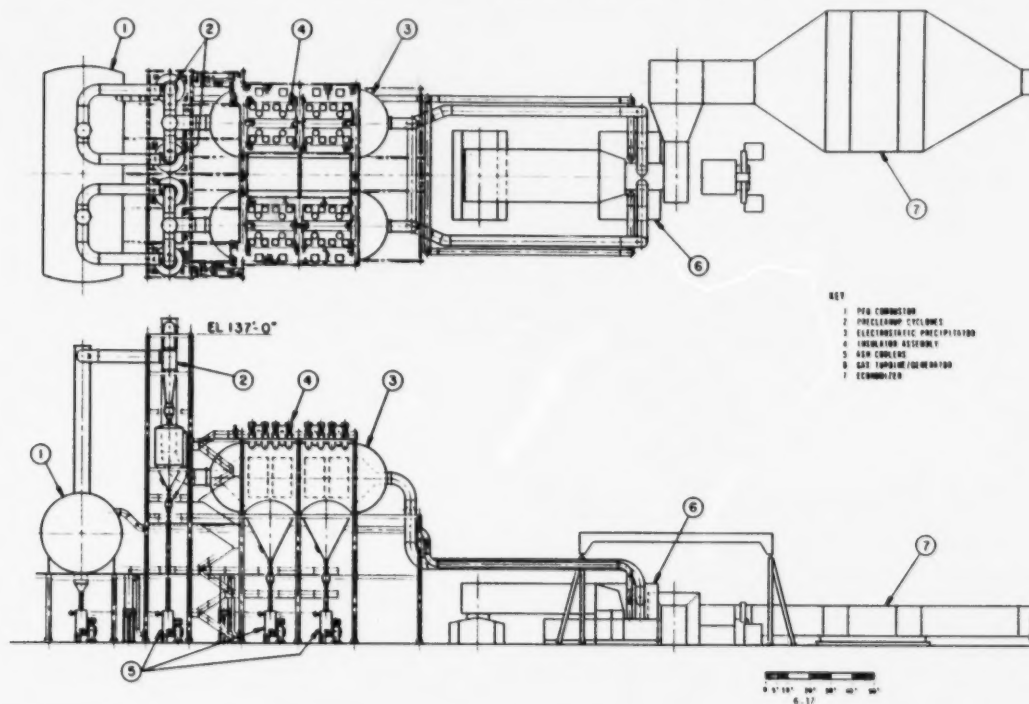


Figure 1 Normalized Arrangement Layout, Electrostatic Precipitator, Cottrell Environmental Sciences.

Maintenance costs were developed with the emphasis on service life and scheduled replacement of severe service components. Even though maintenance estimates for advanced technology cleanup devices are speculative at best, their exclusion could radically distort a comparison between systems which vary greatly in maintenance requirements. Consideration of maintenance costs can be beneficial by drawing attention to areas of needed design development and cost reduction.

Maintenance requirements were identified and consolidated into twelve categories including: refractory, media, high temperature valves, low temperature valves, electrodes, multicyclone internals, advanced device internals, granule transport gas recirculation systems, ash coolers, auxiliary systems, turbines and other components/systems. Also, turbine maintenance penalties were applied to the two plants which, by design, have a high (~200 ppmw) turbine inlet particulate loading, namely, the baseline PFBC and the GE Electrocycle. Annualized maintenance costs were calculated for all categories.

A condensed version of the results is shown in Table 4. Repair or replacement of high temperature ash valves, internals constructed of exotic materials, and refractory lining all figured highly in the maintenance costs.

TABLE 4
HTHP CLEANUP SYSTEMS ANNUAL MAINTENANCE COSTS

(\$ in thousands)

<u>Advanced HGCU System</u>	<u>Total Annual Maintenance Cost</u>
Reference PFBC Plant (Cyclones)	4,963
Electrostatic Precipitator (CES)	6,728
Ceramic Felt Filter (Acurex)	9,342
Ceramic Cross Flow Filter (Westinghouse)	4,862
Shallow GBF (Ducon/Westinghouse)	6,345
Moving GBF (CPC)	4,472
Electrostatic GBF (GE)	4,535
Dry Plate Scrubber (APT)	11,633
Magnetic GBF (Exxon)	5,440
Electrocyclone (GE)	5,066
Acoustic Agglomerator (Aerojet/PSU)	4,635

ECONOMIC ANALYSIS

Normalized capital costs and costs of electricity were developed for the ten HTHP cleanup systems. The HTHP cleanup system capital cost results yielded a cost range of from 117 \$/kW to 293 \$/kW in mid-1983 dollars. Table 5 summarizes these costs. The total cost of electricity values range from 97.0 mills/kWh to 108.3 mills/kWh.

To directly compare major capital cost elements of the subject devices, the estimates were rank ordered as shown in Figure 2. Only the direct cost portion is presented (Table 5 includes indirects). Seven systems compare (close favorably to or less than) to the Baseline PFBC cyclone system, led by the Westinghouse Ceramic Cross Flow Filter; three have greater than a 10 percent advantage. The three HGCU devices shown to be higher than the baseline are decidedly higher by more than 40 percent. Examination of the individual segments reveals a wide variation of primary device costs, with an average cost of \$26.1 million. Due to the wide dispersion in primary device costs, supplementary comparisons were made of device internals costs versus device internal weight, total device cost versus vessel volume, and device ash collection cost versus the number of ash collection points. Correlation in all the investigations was excellent, and the capital costs developed in this study were verified by the comparisons.

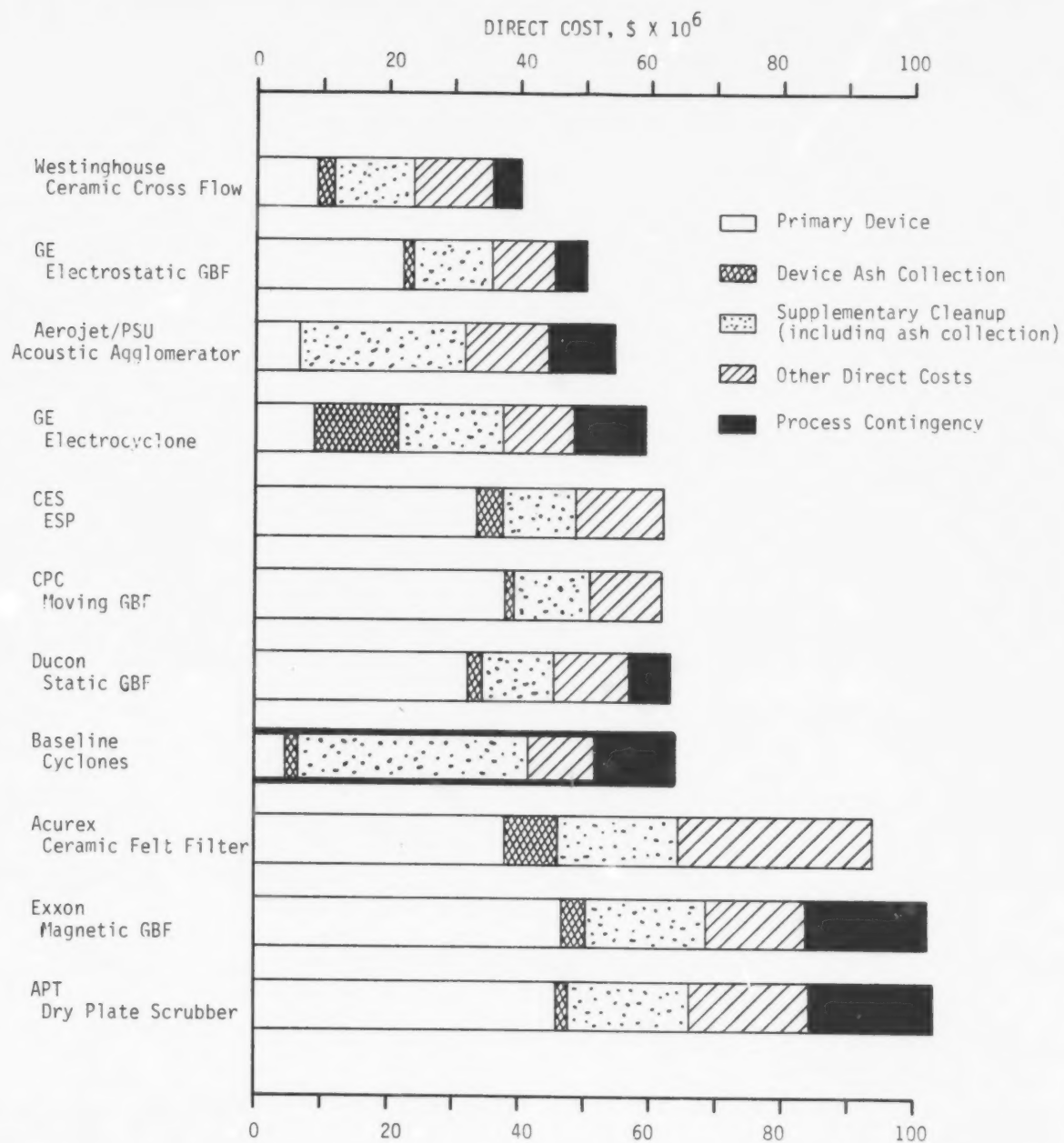


Figure 2 Comparison of Direct Capital Cost Elements.

TABLE 5
PFBC HOT GAS CLEANUP CAPITAL COST SUMMARY
(mid-1983 dollars)

TITLE	TOTAL HGPU SYSTEM CAPITAL COST		TOTAL PLANT CAPITAL COST	
	\$ x 10 ³	(\$/kW)	\$ x 10 ³	(\$/kW)
Conventional Coal Fired Fossil Plant	-		866,900	(1294)
Baseline PFBC Multicyclone	124,200	(188)	795,016	(1200)
Electrostatic Precipitator (CES)	\$125,500	(189)	\$796,254	(1202)
Ceramic Felt Filter (Acurex Corp.)	190,400	(288)	861,192	(1303)
Ceramic Cross Flow Filter (Westinghouse)	78,200	(117)	748,997	(1117)
Shallow Static GBF (Ducon/Westinghouse)	120,300	(180)	791,054	(1185)
Moving GBF (CPC)	120,500	(181)	791,306	(1188)
Electrostatic GBF (GE)	97,100	(146)	767,892	(1153)
Dry Plate Scrubber (APT)	191,500	(292)	862,342	(1315)
Magnetic GBF (Exxon)	189,100	(293)	859,857	(1331)
Electrocyclone (GE)	116,200	(176)	786,963	(1193)
Acoustic Agglomerator (Aerojet/PSU)	104,400	(155)	775,158	(1154)

The accuracy of the absolute value of HGPU system capital costs is probably ± 30 percent, but for comparison purposes the normalized costs are likely to be within ± 10 or ± 15 percent on a relative basis. When the cost breakdown shown in Figure 2 is considered, it is clear that changes to any single aspect of the design must be significant to have an effect upon the system cost. However, certain aspects of the categories shown are directly related to the number of modules required for each device; therefore, a change in this parameter would have an impact in several cost categories. It is also apparent that the seven systems which compare favorably with the baseline cyclone design all appear to be viable in terms of capital cost.

An economic analysis was performed for the ten normalized HTHP cleanup system designs and the reference design. Included in the analysis are separate inputs for capital requirements and fuel consumption as well as operation and maintenance (O&M) costs. The total COE values range from 96.0 to 108.3 mills/kWh as shown in Table 6. Generally, the ranking of COE values follows the capital cost trends; however, the influence of maintenance and fuel cost differences can be seen.

TABLE 6
COST OF ELECTRICITY SUMMARY
(Costs in mills/kWh - 1983 dollars)

<u>Plant</u>	<u>COE</u>	<u>% Change from Base</u>
Conventional Coal Fired	105	+6
Baseline PFBC Multicyclone	99	0
CES Electrostatic Precipitator	101	+2
Acurex Ceramic Felt Filter	105	+6
Westinghouse Ceramic Cross Flow	96	-3
Ducon/Westinghouse Shallow Static GBF	99	0
CPC Moving Bed GBF	98	-1
GE Electrostatic GBF	98	-1
APT Dry Plate Scrubber	108	+9
Exxon Magnetic GBF	107	+8
GE Electrocyclone	100	+1
Aerojet/PSU Acoustic Agglomerator	98	-1

CAPITAL COST TRADE-OFF STUDIES

The need for capital cost trade-off studies for commercial size cleanup devices is apparent when the relative amount of testing that has been done is considered. The HGCU technology is new; the conditions at HTHP are extreme; and as a result, the projected designs and the resulting capital costs must be viewed as preliminary. By varying the fundamental design parameters, the effects on costs, although approximate, can show how optimizing performance can influence cost. They also demonstrate, in some cases, the effect on cost of performance which has been overstated. Examining the effect of increased system operating pressure within the range being considered by PFBC developers provides insight into the relative cost impact of this change. Investigating different HGCU device packaging arrangements provides a check on the arrangement chosen for the normalized design. Figure 3 summarizes the capital cost results of the numerous trade-off studies performed.

The results of the trade-off studies provide an indication of HGCU device cost savings potential and indications of what development areas may have the greatest payoff. The trade-offs do not provide a probability estimate of

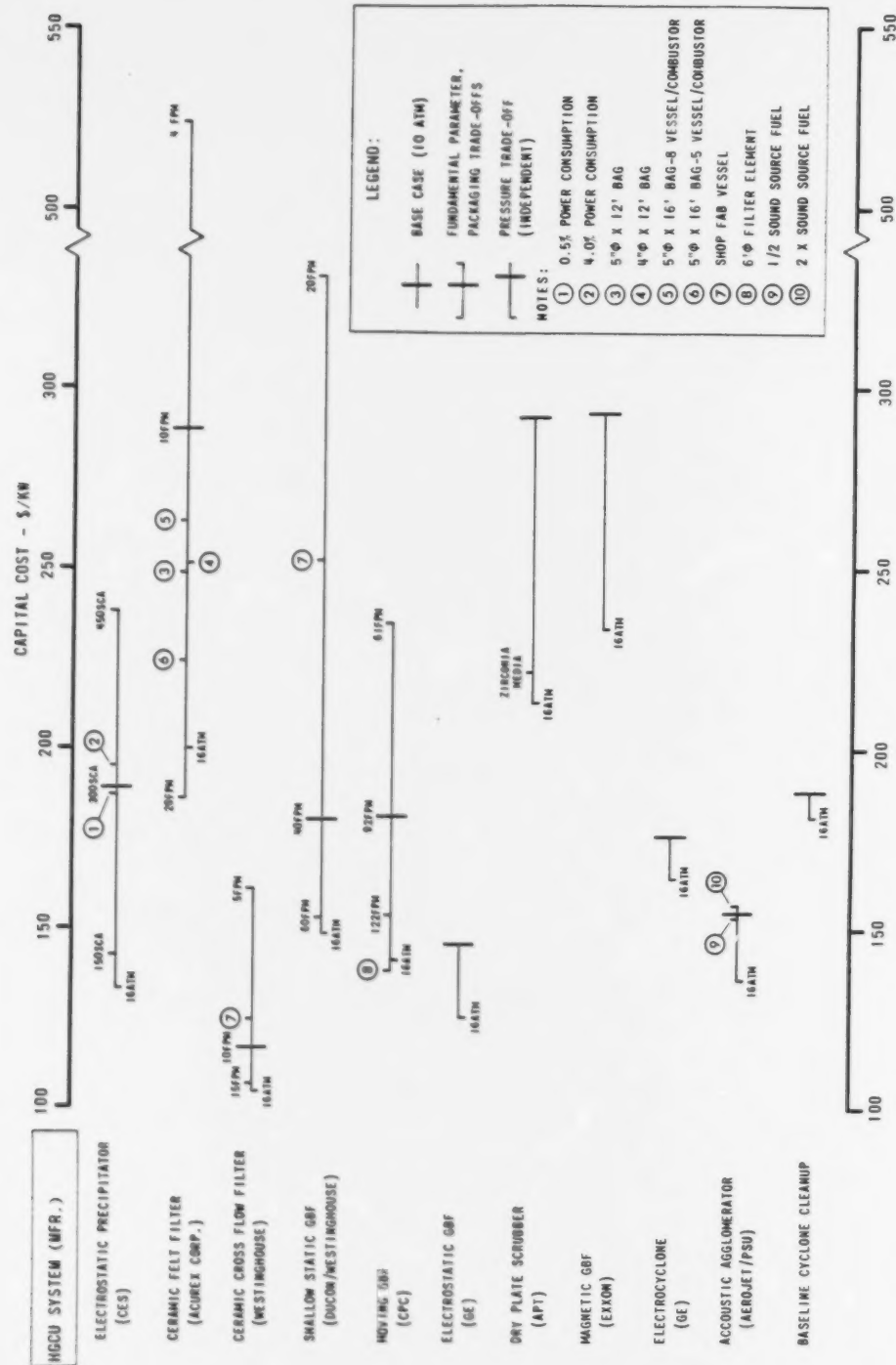


Figure 3 Capital Costs Tradeoff Summary.

achieving any of the stated trade-off ranges; thus it should not be assumed that it is a simple matter to move along the lines shown in Figure 3 for fundamental parameter trade-offs; and comparisons of relative potentials between devices should be done with extreme caution. The results of pressure trade-off studies indicate that a significant capital cost advantage may exist when cycle pressure is increased from 10 to 16 atmospheres. The remainder of the trade-offs showed that significant reductions in capital cost can be achieved by improving the fundamental parameter of the device.

HTHP CLEANUP SYSTEM EVALUATION

Based on the testing completed to date, the commercial size design work performed by vendors of HTHP cleanup systems, the design normalization effort, and the cost analysis of the normalized designs, an evaluation was performed in an attempt to put the development progress and the potential commercial viability of each device into perspective. The decision to proceed with further development of a device/system must be based on as many objective rating parameters as possible, but also, of necessity, on the subjective opinions of experts. The opinions of experts were solicited relating to a variety of parameters which are felt to be crucial to the success of any HTHP particulate cleanup device. Engineers at G/C, DOE personnel, independent experts in HCCU fundamentals, and the contractors themselves participated in the evaluation. The evaluation was considered a success and the correlation of response by individuals and groups enhances the value of the ranking which resulted.

Details of the survey, the results and how the results were used to determine probability of success for the devices is not within the scope of this paper but are available in the G/C report to DOE¹.

CURRENT TEST STATUS

DOE has funded the testing of three devices at the NYU Antonio Ferri Laboratory, Westbury, N.Y. The facility has a 30 inch I.D. PFBC capable of operating near 10 atmospheres with a flue gas volume up to 1800 ACFM at 1600° F. The total gas flow will go thru the cleanup devices. Table 7 lists the devices and the design conditions. Although the test duration, 200 hours for each device, will be short, the data should provide valuable operational information.

G/C was responsible for the design of the test facility, including the structure to support the test device, the process pipe, the particulate collection system and the electrical wiring. Figure 4 shows the CES ESP as installed at NYU. Testing should begin in March, 1986. The most important data collected will be particulate loadings in an out of the devices for efficiency calculations and to determine if NSPS outlet loadings can consistently be met. Grab samples and impactor measurements will be taken periodically and a transmissometer opacity meter will monitor the stack gas continuously. DOE is also continuing the development of the acoustic agglomerator concept and ceramic bag filters but on a lesser scale.

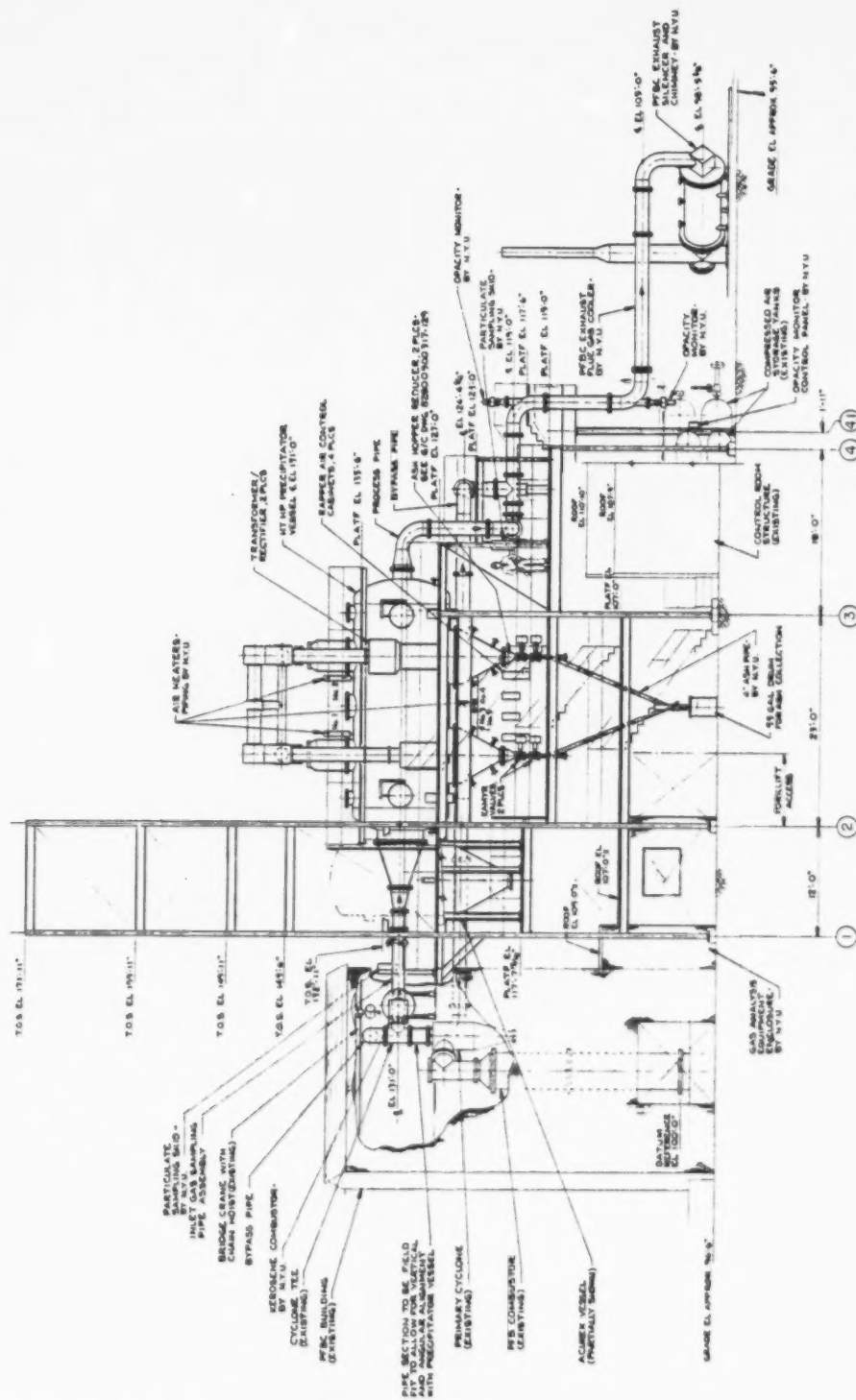


Figure 4: ESP Test Installation At NYU

TABLE 7

DOE/NYU HOT GAS CLEANUP TEST PROGRAM OPERATING PARAMETERS AT10 atmosphere and 1650°F

	<u>1200 ACFM</u>	<u>1800 ACFM</u>
Cottrell Environmental Sciences ESP, Specific collection area	180 SCA	125 SCA
Westinghouse Ceramic Cross Flow Filter Face velocity	8 fpm	12 fpm
Combustion Power Co. Moving Granular Bed Filter Face velocity	60 fpm	90 fpm

EPRI's current interest in HTHP particulate cleanup is in the 800 to 1000°F range, although significant testing has been conducted at higher temperatures. Primary focus has been with barrier type filters, but EPRI also holds promise for the ESP, and has employed this concept in their baseline Turbocharged PFBC design. The two barrier type filters selected for development are woven ceramic bags and porous rigid ceramic filters known as "candles." The "candles" are actually hollow cylinders, closed at one end. The flue gas flows from outside the vertically oriented cylinder to the inside, and up through a tube sheet. Pulse air for cleaning is the reverse direction. Previous DOE sponsored programs and EPRI tests revealed that development work was needed to develop a bag material and configuration that could withstand not only the temperature, but the repeated pulsing required to clean. The current EPRI program has addressed these issues by instituting improved manufacturing methods, design, and testing for bag filters. The 3M Company has been the key developer of the woven bag material. "Candle" filters constructed of porous ceramics are now commercially manufactured for high temperature, low pressure applications. For the 10 atmosphere PFBC environment, enclosure in a pressure vessel is necessary, thus the filter cost is somewhat dependent upon the "candle" length that can be manufactured, and still remain durable, so that an efficient packaging arrangement can be obtained.

A significant number of small scale tests have been conducted to verify that both the bag filter and the ceramic candle filter show considerable promise for 1600° F, 10 atmosphere PFBC applications. Two unanswered questions remain:

1. How do the filters operate in actual PFBC environment?
2. Will the filters operate with acceptable cleanability, dust cake shedding, and chemical degradation without blinding for the long term?

To answer these questions, EPRI is conducting or initiating three test programs.

At the University of North Dakota, a slipstream from the freeboard of the lignite fired stoker-type boiler used for facility steam production. Dust is injected into the 900° F flue gas to match anticipated PFBC loadings and the dust-laden flue gas is passed through a filter vessel containing nine bags. Over 1000 hours were logged in 1985, 5000 hours are expected by June, 1986 and a project goal of 6000 hours should be completed shortly thereafter.

The University of Aachen in West Germany has installed a pressurized fluidized bed boiler to increase their steam output. This unique unit can operate at pressures ranging from one to four atmospheres. A slipstream of approximately 0.4 kg/sec at 1500° F, four atmospheres is passed through a small vessel containing six ceramic "candles." The hardware has been built and is now being installed. Parameteric tests as well as long term durability tests are planned.

The third program is a larger scale test at the National Coal Board/Central Electric Generating Board PFBC test facility at Grimethorpe, England. The 10 atmosphere, 1500° F, test will utilize a 50 percent slipstream, or 7 kg/sec of actual PFBC exhaust. This will be the largest known HTHP particulate cleanup test. The 8 foot diameter 30 foot tall filter vessel is designed to house 130 ceramic "candles" or 44 ceramic bags. An endurance test is not planned, but 1000 to 1500 hours are scheduled as "piggyback" to other planned tests. A decision will then be made to test ceramic bags. Testing should commence early in 1987.(3)

The EPRI program dovetails well into the DOE testing of granular beds, ESP's, and ceramic cross flow filters discussed previously. Promising results have been obtained to date, and the programs in place should provide data from which commercial units can be designed.

Promising demonstration size units could be available for use in PFB Combined Cycle demonstration plants by the year 1988 if EPRI sponsored tests at Grimethorpe are successful. DOE-sponsored test devices could follow in 1989. For this to happen, DOE sponsored tests at NYU would have to be completed in 1986, followed by successful longer term durability tests, preferably at the same facility. Larger scale tests conducted at the Grimethorpe PFBC should be sufficient to support a larger 50 to 100 MWE demonstration plant design if augmented by smaller scale, long-term testing.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

1. Rubow, L. N. et al, "Technical and Economic Evaluation of Ten HTHP Particulate Cleanup Systems for PFBC," Gilbert/Commonwealth Report 2526, U.S. DOE Report No. DOE/MC/19196-1654, July, 1984.

2. Rubow, L. N. et al, "Results of Technical and Economic Evaluation of Ten High Temperature High Pressure Particulate Cleanup System for PFBC," Presented at the Fifth Annual Contractors Meeting on Contaminant Control in Hot Cool-Derived Gas Streams, Morgantown, WV, May 8, 1985.
3. Telcon L. Rubow to S. Drenker, Electric Power Research Institute regarding the EPRI HTHP Particulate Cleanup Program, January 27, 1986.

PARTICULATE CONTROL AT HIGH TEMPERATURE AND PRESSURE FOR EMERGING
ELECTRIC POWER GENERATION TECHNOLOGIES

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ABSTRACT

Particulate control is required to protect turbine components and for environmental reasons in emerging energy technologies, such as Pressurized Fluidized-Bed Combustion (PFBC) and Integrated Gasification Combined-Cycles (IGCC). The Department of Energy (DOE) Fossil Energy Program is developing hot gas particulate control devices for these energy systems. This paper discusses part of the past experimental work by DOE contractors that was the selection basis for bench-scale and subpilot-scale testing of specific hot gas cleanup devices. The evaluation of ten hot gas cleanup devices for use in FBC systems at above 1,000°F and more than 6 atmospheres is described. The suitability of these ten particle control devices for three generic gasifier conditions (fixed bed, fluidized bed, and entrained flow) is also discussed. The current DOE test program is presented along with preliminary results from the bench-scale testing on a fluidized-bed gasifier.

INTRODUCTION

The Department of Energy (DOE) Morgantown Energy Technology Center is sponsoring research to develop coal-fueled power generation systems, which can be strong candidates to supply electric power to the United States in the next several decades. Two emerging energy systems that are of particular importance are the pressurized fluid-bed combustor (PFBC) and the integrated gasifier combined cycle (IGCC). Figure 1 shows a general schematic of a PFBC system. A generic IGCC system is diagramed in Figure 2. Both the PFBC and the IGCC electric power generation systems are more efficient in converting coal to electricity than conventional pulverized coal-fired power plants. The PFBC plant has the potential of reaching 40 percent conversion efficiency and the IGCC has the potential of reaching 45 to 50 percent conversion efficiencies. If the capital costs, environmental performance, and reliability of these systems are realized, they become promising candidates for this country's future electric generating capability.

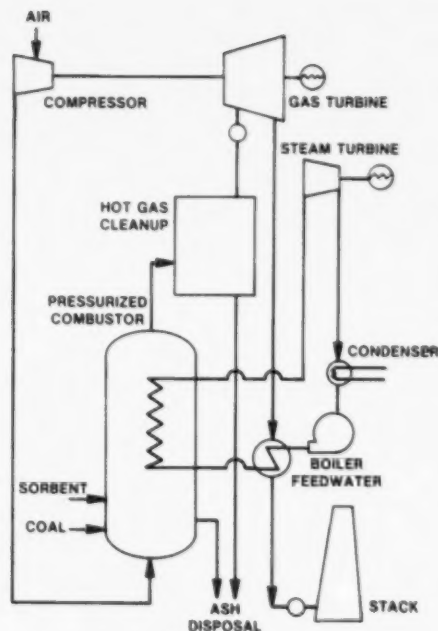


Figure 1
PFBC Steam Cooled System

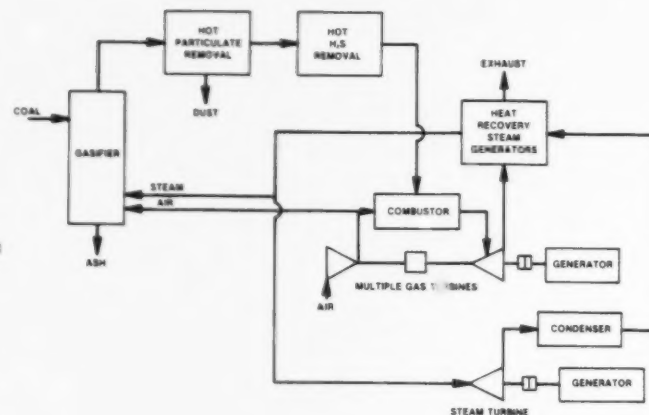


Figure 2
Generic Integrated Gasifier
Combined Cycle (IGCC)

A key element of these systems is the development of particulate control devices to protect the turbines in these energy systems. In addition, these particulate control devices must reduce emissions to a sufficiently low level to prevent any adverse environmental consequences. The PFBC and IGCC systems produce dirty gas streams that require cleaning at temperatures from 1,000° to 1,700°F, and pressures ranging from 6 to 30 atmospheres. The particulate control devices developing under DOE's Hot Gas Cleanup Program are being designed to operate at these conditions.

A primary reason for particulate control in these power systems is to protect the turbine blades that are exposed to the hot/dirty gas from corrosion, deposition, and erosion that could cause significant problems in these turbine systems. Erosion is caused by particles traveling at high speeds impacting on the turbine blades. Deposition occurs when certain contaminants in the fuel deposit on the surface of the turbine's blades and other turbine components. These deposits can build up to form slag layers, thus, impeding gas flows, plugging cooling air circuits, or can provide sites for corrosion. The corrosion is a chemical attack that weakens the materials used in making the turbine components and, in these environments, is principally caused by alkali metal compounds.

The corrosion, deposition, and erosion failure mechanisms of the turbines may be reduced partially or in whole by removing the particulate from the gas flow before it enters the turbine expander. Past research on particulates and turbine components has indicated that it is more important to control larger particulates than smaller particles. As far as the turbine expander is concerned, smaller particles are not as important as larger

particles since particles in the less than 5 microns size range are able to stay in the flow stream and, thus, do not impact on the turbine blade's surface. It is currently felt that a particulate standard of 0.01 grain per standard cubic foot for those particles less than 5 microns and zero particles of greater than 5 microns will provide sufficient protection of the turbine. However, it should be noted that the Environmental Protection Agency is currently considering changing particulate standards to control less than 10 micron particles because of research that indicates that less than 3 micron particles lodge in the human respiratory tract. Therefore, it is possible that in the near future any of these energy systems will have to control small particulates (less than 5 microns) for environmental reasons as opposed to turbine wear reasons.

The reasons for hot gas particulate control devices are to reduce turbine problems associated with particulates and to protect the environment. There are some distinct advantages in performing both functions in the same device at high temperatures and pressure. This advantage can be seen when a single hot gas cleanup device is compared to a conventional PFBC or IGCC power plant design that incorporates hot cyclones to provide some turbine protection, and then has additional particulate control downstream of the turbine (i.e., bag filters) to allow the power plant to meet environmental standards.

HOT GAS CLEANUP STUDIES

Ten particulate control devices were compared to determine the economic and technical viability of each device when compared to a base line PFBC power generation plant. These results were then evaluated to determine the feasibility of applying these ten particulate control devices in gasifier environments. The control devices and the research organization that performed the development work are listed in Table 1.

Each of these control devices has had a very different history in its development. As such, each device has a different engineering data base upon which to judge the developmental progress and the potential for commercial viability.

Gilbert/Commonwealth conducted a study for the U.S. DOE METC to evaluate the technical and economic merits of these ten hot gas particle control devices when applied to PFBC power generation systems (1). This study included a ranking of the ten devices by 24 experts in the hot gas cleanup field as well as objective rankings by economic criteria. These experts included Gilbert/Commonwealth engineers, DOE technical personnel, independent experts and consultants, and the individual contractors that have developed the high-temperature, high-pressure (HTHP) cleanup system. Biases in the evaluations were minimized by the group selections, and any possible biases by the individual contractors were offsetting.

It is felt that the experts in the hot gas cleanup field are best able to identify a variety of parameters that could be crucial to the success of any HTHP particulate control device. The opinions of these experts were formulated into an analysis which weighed the risks versus the potential benefits

TABLE 1. Ten Particulate Control Devices Evaluated
in the DOE Hot Gas Cleanup Program

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1. High-Temperature/High-Pressure Electrostatic Precipitator developed by Cottrell Environmental Sciences -- CES ESP
 2. Ceramic Felt Filter developed by Acurex Corporation -- Acurex Filter
 3. Ceramic Cross-Flow Filter developed by Westinghouse Electric Corporation -- Cross Flow Filter
 4. Shallow Static Granular-Bed Filter developed by Ducon and Westinghouse Electric Corporation -- Ducon/Westinghouse
 5. Electrostatic Granular-Bed Filter developed by General Electric Corporation -- GE EGBF
 6. Moving Granular-Bed Filter developed by Combustion Power Corporation -- CPC Moving GBF
 7. Dry Plate Scrubber developed by Air Pollution Technology -- APT DPS
 8. Magnetic Granular-Bed Filter developed by Exxon Corporation -- Exxon MGBF
 9. Electrocyclone developed by General Electric Corporation --Electrocyclone
 10. Acoustic Agglomeration developed by Aerojet Energy Conversion Corporation/Pennsylvania State University -- Aerojet/PSU AA
-

of further research on each particulate control concept. Thus, the relative ranks for each device can be used as a risk analysis that would quantify the likelihood of reaching the DOE Hot Gas Cleanup Program goals.

The survey and evaluation was concerned with three main areas: basic concepts testing, commercial viability, and probability of success. The probability of success area included the likelihood of meeting new source performance standards (NSPS), achieving acceptable cost, and the chances of being able to have sufficient reliability. Table 2 shows the probabilities that were assigned to each rating and applied by the respondents.

In addition to assigning probabilities to the success of the ten control devices, each respondent was asked to select the three devices that they felt should be further developed based upon the device's potential and probability of success.

To build upon the evaluation of particle control devices for PFBC systems, a second study was performed to evaluate the applicability of these hot gas cleanup devices (HGCU) for use in coal gasification systems (2). This study was performed by establishing the gas stream characteristics that are important for the collection of particulates. For each of the ten hot gas cleanup devices, a detailed assessment of the projected particle collection performance was conducted, along with the design and operating parameters. Each of the hot gas cleanup devices that were originally designed for PFBC application were then ranked according to their suitability for use in coal gasification systems. Three main issues were raised when a given HGCU

TABLE 2. Probabilities of Success Applied to Each Rating Level

Rating	Probability of Success (Percent)
0	< 5
1	5 - 25
2	26 - 50
3	51 - 75
4	76 - 95
5	> 95

device was considered for its effectiveness under coal gasification conditions. The first issue was if the device could be reasonably expected to meet the perceived NSPS standard for particulates. Secondly, would the physical nature and chemical composition of any contaminants (i.e., alkali compounds) that escaped the HGCU device endanger the downstream turbine or other equipment. The third issue for considering a HGCU device was whether the device could be expected to have a reasonably long service life and operate without an excessive energy penalty in a coal gasification environment.

Due to the fact that these hot gas cleanup devices were principally being developed for PFBC applications, there was no body of test data that could be used directly to predict the performance of the devices under coal gasification conditions. Further, there is very limited data on the behavior and nature of gas streams and particulates in a combined-cycle system that integrates turbines and gasifiers. Therefore, a set of five fundamental collection parameters describing the physics of particulate collection were used to model which gas stream variables (i.e., temperature, pressure, gas flow rate, particle diameter, particle loading, carbon content, alkali loading, and tar loadings) would most strongly influence particulate collection. These fundamental parameters include inertial impaction, direct interception, gravitational settling, diffusion, and electrostatic collection.

The model predictions were emphasized for the prediction of each device's particulate collection efficiencies due to the lack of test data for these devices under gasifier conditions. Three sets of generic gasifier conditions were used to evaluate the hot gas cleanup devices. These conditions are listed in Table 3.

The study conditions listed in Table 3 were chosen to represent exit gas stream characteristics from a cross-section of coal gasification processes, and at the same time to cover the extremes which the HGCU devices might face. The study compiled data from commercial and near-commercial gasification systems to evaluate what the study conditions would be. Since this study was intended to evaluate particulate control devices in all types of

TABLE 3. Study Conditions for Evaluation of HGCU Devices for Coal Gasifiers

	Condition 1 Fixed Bed	Condition 2 Fluidized Bed	Condition 3 Entrained Flow
Temperature (°F)	800	1,200	1,600
Pressure (atm)	1	5	20
Particulate Loading (mg/m ₃)	600	9,000	9,000
Mean Particle Size (micron)	2	25	100
Particle Size Standard Deviation (micron)	2.4	3	3

gasifier systems, the conditions listed in Table 3 are not considered the optimum conditions that a gasifier would be operated while integrated in an IGCC power plant. This drawback should not change the study results to any significant degree primarily because this study has pointed out that the dominant variable for determining particulate collection is particle diameter. Gas flow and temperature were the next most important variables, and pressure was found to be insignificant. Thus, it can be concluded that for condition 1 -- fixed-bed gasifier, the study results would not change appreciably for a pressurized gasifier that operated at a somewhat higher temperature. One of the major considerations in evaluating HGCU devices for the fixed-bed gasifier is the potential for tars and the concern that these tars would blind the filtration type devices.

The combination of these studies along with important conclusions from other DOE Fossil Energy Programs have influenced the selection of particle control concepts for continued research and development. The Fluidized-Bed Combustion, Gasification, and Heat Engines Programs being managed at DOE METC have provided invaluable data and design specifications for hot gas cleanup systems. Of particular importance are results that were developed under the Heat Engines program where minimally cleaned coal gas was combusted in conjunction with a turbine simulator. It has been important to evaluate the potential of each device for both gasifier and PFBC systems when test programs are laid out.

RESULTS

Table 4 shows the ratings that each device was given for the probability of meeting NSPS in a PFBC power system, achieving acceptable costs, and meeting acceptable life/reliability criteria.

The values under the composite rating column are the average of the other three rating columns. These ratings are the average of all five groups and include the responses of each hot gas cleanup device's developer. The developers' responses did not bias the results to any degree. The results of the composite rating column in Table 4 shows that an ESP, a moving

TABLE 4. Probability Ratings for HGCU Development

Device	Composite Rating	NSPS Rating	Cost Rating	Reliability Rating
Cottrell ESP	4.0	4.3	3.7	4.1
CPC Moving Granular-Bed Filter	3.7	3.8	3.7	3.7
Cross Flow Filter	3.6	4.1	3.8	3.0
Acurex Filter	3.1	4.1	2.7	2.5
GE Electrostatic Granular	3.0	3.1	3.2	2.8
Aerojet/PSU AA	3.0	2.4	3.6	3.2
Electrocyclone	2.7	1.5	3.5	3.1
Ducon/Westinghouse	2.7	2.6	3.0	2.6
Exxon MGBF	2.0	2.5	1.2	2.2
APT DPS	1.9	2.3	1.6	1.9

granular-bed filter, and a cross-flow filter are the three highest rated devices. Table 4 also shows that the these three devices demonstrate a good probability of meeting NSPS, with the ESP given the highest rating, and the moving granular-bed filter given a slightly lower rating. The cross-flow filter is rated highest of all ten control devices for achieving acceptable costs. For achieving reliability Table 4 shows that the ESP is rated as the most reliable, followed by the moving GBF. There were several other devices rated more reliable than the cross-flow filter, which primarily reflected concerns about sealing and cleaning the filter elements.

Each of the respondents were asked to recommend three devices for continued development based on the cleanup device's potential and probability of success, not development progress alone. Table 5 shows these results.

It should be noted that the results in Table 5 also are not changed by including the recommendations by the contractors that have developed these cleanup devices. The results in Table 5 show that three devices that had the largest number of recommendations were the same devices that had the highest composite ratings in Table 4.

Table 6 combines the major results from Tables 4 and 5 with the results from Gilbert/Commonwealth's economic analysis. When the ratings in Column 1 and the recommendations in Column 2 of Table 6 are compared to the capital cost in dollars per kilowatt for each hot gas cleanup device, it is evident that there are a number of devices that do not warrant further research. In general, these devices are the bottom five devices listed in Table 6. DOE is concentrating the Hot Gas Cleanup Program on the top five devices in Table 6, these being ESP's, moving granular-bed filter, cross-flow filter, ceramic bag filters, and acoustic agglomeration.

It should be noted that the systems direct capital cost for the Acurex filter systems is partially due to the many high-temperature valves that were designed into their cleanup system. It has been felt that by redesigning the configuration of the filters or by improvements in high-temperature

TABLE 5. Recommendations for Continued Development

Device	Total	Number of Recommendations				
		G/C HGCU Engineers	G/C Power Engineers	DOE METC	Consul- tants	Contrac- tors
CES ESP	21	5	3	5	3	5
CPC Moving GBF	14	4	2	2	3	3
Cross Flow Filter	13	3	2	4	0	4
Acurex Filter	9	0	1	2	2	4
Aerojet/PSU AA	6	2	0	2	0	2
Electrocyclone	2	1	1	0	0	0
GE EGBF	2	0	0	0	1	1
Ducon/Westinghouse	1	0	0	0	0	1
APT DPS	0	0	0	0	0	0
Exxon MGBF	0	0	0	0	0	0

TABLE 6. Comparison of Expert Opinions, Recommendations, and Capital Costs for 10 HTHP Particulate Control Devices on PFBC

Device	Composite Ratings	Recommen- dations	Total System Direct Capital \$/kW
CES ESP	4.0	21	93
CPC Moving GBF	3.7	14	94
Cross Flow Filter	3.6	13	54
Acurex Filter	3.1	9	142
Aerojet/PSU AA	3.0	6	67
Electrocyclone	2.7	2	73
GE Electrostatic GBF	3.0	2	68
Ducon/Westinghouse GBF	2.7	1	85
APT Dry Plate Scrubber	1.9	0	129
Exxon Magnetic GBF	2.0	0	130

valves which reduce valve costs, the Acurex filter economics would become significantly better.

These results reflect a major part of the information used in selecting concepts for continuing R&D but, as was mentioned earlier, there has also been a study of these ten hot gas cleanup devices for gasification systems. The results of this gasifier cleanup study indicated which of the HGCU devices would be suitable for the three gasifier conditions that were listed in Table 3. The results are presented in Table 7.

The electrostatic precipitator (ESP), moving granular-bed filter, and acoustic agglomeration are the only three devices that were evaluated as being likely cleanup devices for fixed-bed gasifier conditions. This is primarily due to the tar content that is typical of fixed-bed gasifier gas streams.

TABLE 7. Hot Gas Cleanup Devices Evaluated as Suitable for Gasifier Systems

Condition 1 Fixed Bed	Condition 2 Fluidized Bed	Condition 3 Entrained Flow
ESP Moving GBF Acoustic Agglomeration	ESP Moving GBF Acoustic Agglomeration Ducon/Westinghouse GBF Acurex Bag Filter Cross-Flow Filter	Cross-Flow Filter Moving GBF Acoustic Agglomeration Ducon/Westinghouse GBF Acurex Bag Filter ESP

Some reasons behind these choices are that ESP's are suited to tar laden gas streams because they do not blind, GBF's have a high thermal mass which is advantageous if gas stream temperatures drop, and acoustic agglomeration should not be reduced if there are tars present. The three top devices that are listed under Condition 2 Fluidized Bed, were ranked higher than the lower three devices. The main reason for lower rankings of the Ducon/Westinghouse GBF, Acurex bag filter, and cross-flow filter under Condition 2 is because of concern regarding the ability to backflush these devices in order to clean them adequately. The three devices that were ranked as most likely to meet the Condition 3 Entrained Flow environment were the moving granular-bed filter, cross-flow filter, and acoustic agglomeration. The bottom three devices listed under the Condition 3 Entrained Flow column of Table 7 were ranked as less likely candidates than the top three devices.

Examination of the three study conditions showed that for each condition (i.e., fixed bed, fluidized bed, and entrained flow) there are at least three hot gas cleanup devices that were found to be suitable without any fundamental barriers to their potential effectiveness. Two of the devices, the moving granular-bed filter and acoustic agglomeration, were judged suitable for all three coal gasification conditions. There is more confidence in this judgment for the moving granular-bed filter than for acoustic agglomeration. This is because the possibility has been raised that high temperature conditions attenuate acoustic energy and, therefore, compromise the performance of acoustic agglomeration in gasification systems. Thus, the conclusion that acoustic agglomeration is suitable to all three coal gasification systems should be considered tentative.

The modeling work that was a large basis of this study showed that particulate diameter was the dominant variable for all collection mechanisms. Gas flow and temperature were determined to be the next two most important variables for particulate collection in these HGCU devices. Under gasification conditions an important parameter is the formation of tar fogs, which is linked to particle diameter and temperature. The system pressure at which these hot gas cleanup devices are studied was found to be insignificant in affecting the particulate collection mechanisms, however, it should be noted that system pressure is very significant in determining the overall economics of the cleanup system.

Work has also been conducted at the General Electric Corporate Research and Development Center on the effect of hot gas streams on turbine components (3). Part of this work investigated the problems associated with the combustion of hot minimally cleaned low-Btu fuel gas from a coal gasifier in a simulated gas turbine environment. The test apparatus is shown in Figure 3. A pressurized fixed-bed gasifier was used in these tests and provided a hot fuel gas to the combustor, an air-cooled turbine nozzle sector, and ash deposition specimens located downstream of the nozzle. The impingement pins were used to measure ash deposition after the raw fuel gas had been burned in the combustor. One cyclone was used for hot gas cleanup of the gas stream before the combustor simulator. The gasifier was air blown and operated during these tests on Illinois No. 6 coal. Two tests were conducted for a total of 142 hours.

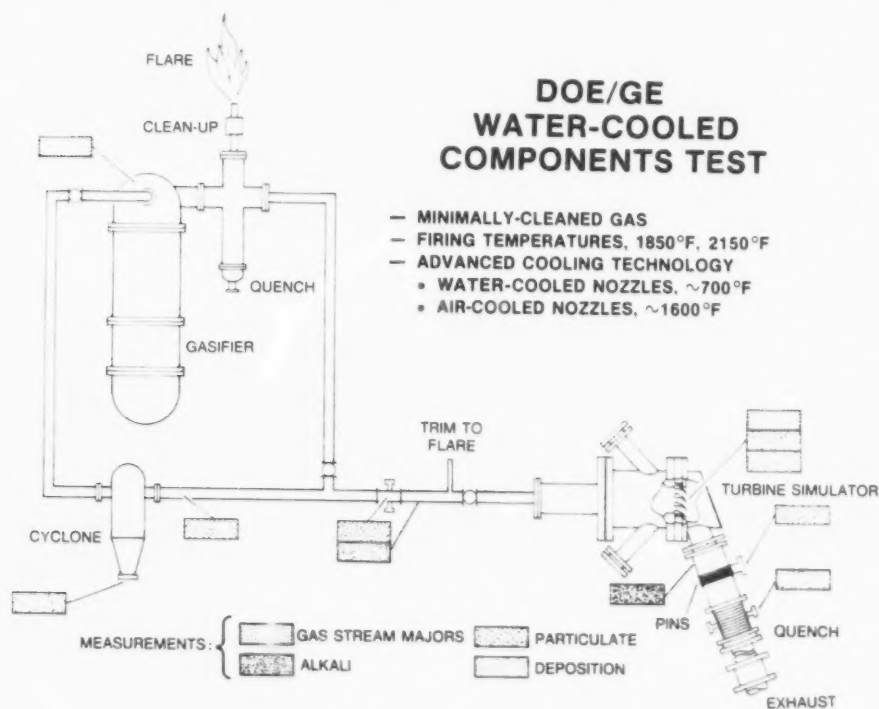


Figure 3
GE Air-Cooled Components Test Facility

The results from this work showed that an average of 46.6 lb/hr of dust exited the gasifier to the cyclone. The cyclone averaged 98 percent collection efficiency which resulted in an average of 0.887 lb/hr (104 ppm) of dust to the turbine simulator. The ash content of the dust was 11.39 percent, which resulted in only 11 to 13 ppm of ash entering the turbine simulator. Very little of this ash was deposited on the airfoils of the air-cooled nozzle cascade during the approximately 100-hour tests. The only particulate erosion detected was on the fuel nozzle, and the erosion was reduced to an acceptable level by using Hastelloy X.

The results from the minimally cleaned hot coal gas combustion tests were important because it was evident that in a gasifier/combustor arrangement (as in an IGCC system) the total particulate loading to the combustor is not as critical as the particulate loading is in a PFBC system. This is because the particulate produced by the gasifier will consist primarily of carbon. In the combustor the carbon will be converted and the combustor exhaust will only contain particulates that are composed of ash. Consequently, in the GE minimally cleaned hot coal gas tests, the 100 ppm of dust leaving the cyclone resulted in only 11 to 13 ppm ash entering the turbine simulator with much of the particulate carbon in the cyclone exhaust being burned.

DOE HOT GAS CLEANUP PROGRAM

The results of the studies discussed, along with previous testing that was conducted at Curtiss-Wright (4,5), have had an impact on the scale of testing and the HGCU devices to be tested. An independent review of the DOE Hot Gas Cleanup Program has also provided certain recommendations (6). Therefore, the DOE Hot Gas Cleanup Program has developed two major sets of projects to address these results and recommendations. These two projects are the testing of hot gas cleanup devices on a subpilot-scale PFBC, and the evaluation of several hot gas cleanup devices on a bench-scale coal gasifier. The subpilot-scale PFBC and the bench-scale coal gasifier projects are discussed in greater detail below.

Subpilot-Scale Testing

Three hot gas cleanup devices are being constructed and will be tested at a subpilot-scale PFBC operated by New York University in Westbury, New York. These three devices are:

- Electrostatic precipitator, designed by Cottrell Environmental Sciences.
- Ceramic cross-flow filter, designed by Westinghouse Electric Corporation.
- Moving granular-bed filter, designed by Combustion Power Corporation.

This facility, the design conditions for each control device, and the general test plan have been described in a separate paper (7) and will not be discussed in this paper.

Bench-Scale Testing

A bench-scale, fluidized-bed coal gasifier is being used to test and evaluate three hot gas cleanup devices. The six-inch diameter gasifier is located at the Morgantown Energy Technology Center, Morgantown, West Virginia. The three devices that are currently in the process of being tested are:

- Ceramic cross-flow filter, designed by Westinghouse Electric Corporation.
- Ceramic bag filter, designed by Acurex Corporation.
- Electrostatic precipitator, designed by Denver Research Institute.

As it can be seen from the devices that are currently being tested and evaluated under the DOE Hot Gas Cleanup Program, the top four rated devices in Table 6 are being looked at under either the subpilot-scale PFBC testing or the bench-scale gasifier testing. In addition, the devices that are being evaluated at METC with the bench-scale gasifier are all listed in Table 7 under Condition 2 Fluidized-Bed Gasifier.

Figure 4 shows a schematic of the Hot Gas Particulate Test Facility at Morgantown. This facility has the capability of testing particulate control devices on gasifier fuel gas at pressures from 100 to 300 psig, temperatures from 1,200° to 1,600°F, and dust loadings from 0 to 5,000 ppm. Gas flows could range from 1,000 to 5,000 scfh (0.02 to 0.10 lb/sec). The test facility employs a supplemental gas heater to boost the temperature of the fuel gas up to the test conditions. Particulate loadings and filter collection efficiencies were determined by a probe in the inlet gas stream. The outlet particulate loadings were determined by filtering the entire gas stream that had left the control device test module through one or two thimble filters operated in parallel. The filters used on the inlet probe and outlet sampling system were ceramic alundum extraction thimbles. The ceramic thimble filters were kept at the system pressure, and while sampling the thimbles were kept at 1,000°F to prevent any condensation. Table 8 gives the typical gas composition that was being produced by the six-inch fluidized-bed gasifier operating on a high-sulfur (3.37 percent) Montana Rosebud subbituminous coal.

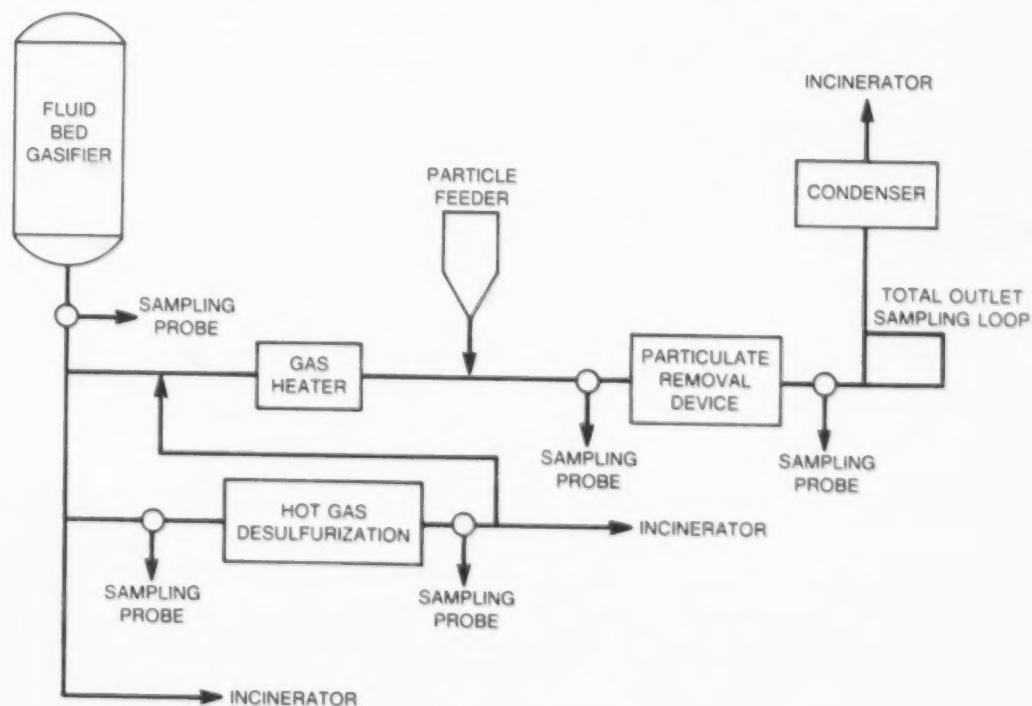


Figure 4
Schematic of Bench-Scale Hot Gas Particulate Test Facility

TABLE 8. Typical Gas Composition for Bench-Scale Gasifier
Particulate Control Device Tests

Component	Volume Percent
H ₂	16%
CO ₂	16.5%
N ₂	43%
CO	8%
CH ₄	3.5
H ₂ S	0.45%
Ar + O ₂	0.55%
H ₂ O	12%

Preliminary results from the bench-scale testing at Morgantown have shown that the Westinghouse ceramic cross-flow filter and the Acurex ceramic bag filter do not have any significant material problems that would preclude using these control devices in gasifier systems. The operation of these filters on fluidized-bed char has definitely shown that this type of char is significantly easier to clean than PFBC dust. This finding means that generally, filtering devices that are developed for PFBC applications will also be applicable for gasifier conditions. It should be noted, however, that this type of gasifier char seemed to penetrate more, thereby causing lower collection efficiencies. It is not certain that all gasifiers will produce particulates that are either easily cleaned or as penetrating. In general, a particulate control device on a gasifier in an IGCC system can be allowed to have lower collection efficiencies since the volume of gas being cleaned per million Btu's is less than under PFBC conditions (no excess air). Also since the particles that do escape the control device are typically 50 to 70 percent carbon, the particles will be substantially burned up in the gas turbine combustor.

Ceramic Cross-Flow Filter

During the testing of the ceramic cross-flow filter there were a number of interruptions; however, because of the flexibility of the Hot Gas Particulate Test Facility, it was usually possible to come back to the same temperature, pressure, and flow conditions. In this way the tests have been combined for each test condition and the particulate loadings averaged over the typical 50-hour time periods that were operated at each test condition. Table 9 shows the average collection efficiency for the cross-flow filter.

The testing of the ceramic cross-flow filter at Morgantown definitely showed that the problems with the filter element delamination had not yet been solved. There were a total of three filter elements tested at Morgantown because the first two elements delaminated during testing. Although the cyclic nature of the testing (caused by the interruptions) could have caused thermal stresses which contributed to the delaminations, it was obvious that some more development work is needed on the filter elements. To address this factor the filter element manufacturers were constantly

TABLE 9. Ceramic Cross-Flow Filter Test Conditions and Average Collection Efficiencies in the METC Tests

Temperature	Pressure	Filter Face Velocity	Inlet Dust Loading	Average Collection Efficiency
1,000°F	250 psi	4.89 ft/min	7,975 ppm	99.47%
1,200°F	250 psi	2.91 ft/min	1,112 ppm	99.25%
1,500°F	250 psi	5.49 ft/min	2,337 ppm	98.44%
1,400°F	150 psi	8.27 ft/min	593 ppm	98.74%
1,200°F	150 psi	8.1 ft/min*	2,408 ppm	98.34%

* Test conducted on hot nitrogen instead of gasifier fuel gas.

appraised of the delamination problems and a series of modifications to the filter elements were tried throughout the Morgantown testing series. The most promising modification consisted of building the filter element with a very thick mounting flange to resist mounting stresses, coating the front face of the filter element with a slurry of the parent ceramic material to increase the ceramic density, and firing the filter element at the highest temperatures allowed by the ceramic material to improve bonding between the ceramic membrane layers. This approach seems to have the most promise in solving the ceramic cross-flow filter element delamination problem, and it should be noted that the last filter tested at Morgantown included these features. This filter was the only element that did not delaminate, yet it was exposed to some of the highest temperatures (1,600°F) for over 100 hours.

Ceramic Bag Filter

To date, the ceramic bag filter has been tested at Morgantown for 110 hours. The filter has operated at 1,200°F and 150 psi at a filter face velocity of approximately 3 feet/minute. The dust loading averaged 265 ppm for one 50-hour test segment and 2,400 ppm for a second 50-hour test segment. During this time period, collection efficiencies started at more than 99 percent, but fell to a 90 percent average collection efficiency during the second test period. An examination of the bag showed that stretching of the filter caused pinholes that eventually allowed a significant amount of the gasifier char to pass through the ceramic bag filter.

CONCLUSIONS

To further the engineering data base for hot gas cleanup the Department of Energy is conducting two sets of projects that will be testing a number of devices on a subpilot-scale PFBC and a bench-scale coal gasifier. These two sets of projects have been specifically chosen to represent the DOE's best estimates of required hot gas cleanup technologies for PFBC and IGCC power systems. The devices that are being tested under PFBC conditions are the devices that were rated as being most likely for success. For the most

part, these devices have been estimated to be some of the lowest cost devices for PFBC applications. In addition, the devices that are being tested under gasifier conditions have been listed as suitable for fluidized-bed gasifier conditions. Although the conditions differ between the PFBC and gasifier testing, both sets of projects will complement each other by addressing certain design or operating problems specific to each hot gas cleanup device. As an example, the testing of a woven ceramic fabric bag filter by Acurex Corporation on the bench-scale coal gasifier will address some of the filter material problems, the results from which can be applied to gas streams other than coal gasifiers. Similarly, since the moving granular-bed filter was judged to be suitable for all three generic types of gasifiers, the test program on the subpilot-scale PFBC will address certain reliability and feasibility issues that can be applied if the moving granular-bed filter is ever tested for particulate control in IGCC systems.

In general, the studies that have been conducted to date indicated that most of the hot gas cleanup devices being evaluated under the DOE program are more economical than a conventional cyclone and post-turbine cleanup system. The results of ongoing work with these hot gas cleanup devices should further improve the reliability and increase the engineering data base needed to put a hot gas cleanup device in commercial-scale PFBC or IGCC power systems.

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THE TESTING AND EVALUATION OF CERAMIC FILTER FABRICS

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ABSTRACT

This paper reports testing methodology and test results of techniques to evaluate ceramic filter fabrics for use in hot gasifier and pressurized fluidized-bed combustor gas streams. Filter media, employing ceramic fibers in their construction for use at temperatures ranging from 600° to 1,600°F, have been under development for several years. Rapid test methods have been developed that can be used to quantitatively assess appropriate properties of ceramic fabrics to be used as hot gas cleaning filters. Key ceramic fabric properties have been established from a consideration of field data and basic mechanisms that affect filter performance. Test methods were then developed for measuring these fabric properties. The usefulness of each test method for measuring ceramic fabric properties has been evaluated by testing different ceramic fabrics under four different conditions of temperature exposure. Evaluation of specific fabrics was done primarily to characterize the test methods and was not intended to determine which ceramic fabrics were better than others. However, testing of ceramic fabrics using different methods provided valuable insight into how ceramic fabrics respond to temperature and external stresses and what fabric properties may be expected to affect filter performance. The methodology developed above was successfully used to prescreen ceramic filter candidates for gasifier applications before evaluation in an actual gasifier stream at the Morgantown Energy Technology Center, Morgantown, West Virginia. Prescreening tests clearly indicated differences among the collection efficiencies of the ceramic fabrics that were tested and indicated that one fabric was preferred over the others.

INTRODUCTION

During the past several decades, fabric filters have been used successfully for high-efficiency particle removal from gas streams. The application temperature is typically limited to less than 550°F because standard filter media, such as Nomex, Gortex, Teflon, glass, and a variety of synthetic and natural fibers deteriorate rapidly at higher temperatures. Promising developments in pressurized fluidized-bed combustion (PFBC) and integrated gasification combined-cycle systems (IGCC) have identified a need for particle control technologies that are operable at temperatures much in excess of 550°F. A variety of fibers comprised of ceramic or metallic materials are now available for use as filtration media at temperatures exceeding 1,000°F. Several types of ceramic filters are currently under development, including woven, matted, and rigid structures. These ceramic filters are made from silica, alumina, zirconia, boria, magnesia, and silicon carbide-based materials with a useful temperature range exceeding 2,000°F.

The United States Department of Energy (DOE) and the Electric Power Research Institute (EPRI) have sponsored tests of ceramic fabrics for high-temperature applications (1,2,3). These tests have illustrated the merits of using ceramic fabric filters for temperatures in excess of 1,000°F, but have also illustrated problems with the technology for specific applications. Mechanical failure of the fabric, seepage of dust through the fabric, fabric embrittlement, failure of sewing threads, and failure of creased seams were observed in the various tests. While many of the problems have been resolved, the results pointed to a need for small-scale tests that can rapidly assess the potential of ceramic fabrics for high-temperature filtration and point out potential problems. This is especially true since many varieties of ceramic materials are potentially suitable for particulate removal, but little information is available on their use for filters. Also, the large-scale, multiple-filter tests are too expensive and time consuming for initial filter screening and would provide more meaningful results after filter performance had been established.

A large number of laboratory tests have been designed to measure the properties of conventional textile fabrics and to evaluate the performance of filter media. Many tests have American Society for Testing and Materials (ASTM) or other standard designations and other tests are special tests used by single filter manufacturers. Various efforts have been made to relate standard test results to actual filter performance (4). Good correlation between standard test results and actual filter performance has not been established for conventional textile fabrics. These standard tests are most important for quality control of fabrics that have proven to be successful in operation.

In the case of ceramic filter fabrics, little, if any, quality control or screening tests have been conducted. It has been shown through studies conducted by Acurex Corporation that tests designed to assess properties of conventional filter media can be modified or extended to evaluate ceramic fabrics (5,6). Newly developed tests that take into account the different

class of materials and different operating environment have been shown to be potentially useful in screening ceramic fabrics for high-temperature filtration. The ultimate usefulness of the screening tests has not been proven and will not be, until some correlation can be established with full-scale operations.

CERAMIC FABRIC TESTING

The observed failures of ceramic fabrics have occurred during a short period of use, generally less than 200 hours (1). In many cases, catastrophic deterioration in mechanical properties such as brittleness, flexibility, and tensile strength can be attributed to temperature and perhaps chemical species in the gas or dust. In the evaluation of filter performance, it becomes important to measure these mechanical properties and develop a profile as a function of exposure time. It is also important to define the type of gaseous conditions and flyash used. The lack of sufficient data precludes a prediction of what minimum values of mechanical properties the filter fabrics need to have in order for the fabric to last a reasonable length of time. In some cases, the filter fabric may not be affected by temperature or chemical species but more by fatigue due to extended use, such as repeated flexing in the cleaning cycle. Although failure due to such a mechanism has not been observed in filters tested so far, longer periods of use may lead to breakage of the filter bags around some high stress areas. Figure 1 shows the area of stress during filter use. Analysis of the stress levels encountered can provide an indication of potential problem areas and estimates of minimum levels of mechanical properties that must be maintained. A consideration of the physical mechanisms and stresses that could affect filter life would help to identify the proper test methods to use for specifying filter fabrics and estimating the expected performance of potential fabrics.

Recent studies conducted by Acurex Corporation for the U.S. DOE Morgantown Energy Technology Center identified data that are useful for characterizing and specifying ceramic filter fabrics for PFBC systems (6). A listing of these data is presented in Table 1. Much of these data can be obtained by using conventional test methods adapted for the special application, but fabric properties such as heat flex, crush resistance, collection efficiency, and cleanability require special equipment for testing. In the first Acurex study, fabric properties were measured under the four conditions delineated below:

- As received
- Heat cleaned at 1,000°F to remove organic sizing
- Heat treated in air at 1,650°F for 100 hours
- Heat treated in air at 1,650°F for 500 hours

Six ceramic fabrics were selected for study and were tested in order to evaluate the usefulness of standard and new test methods for characterizing ceramics for high-temperature filtration. The basic properties of the

fabrics that were tested are delineated in Table 2. It should be noted that both matted and woven fabrics were chosen for testing.

RESULTS OF TEST METHOD CHARACTERIZATION

Results of tests using a combination of standard and new techniques are summarized below. Details of test apparatus and procedures are described elsewhere (5,6).

TENSILE STRENGTH AND MULLEN BURST

Figures 2 and 3 show the results of tensile break strength and Mullen Burst testing of woven ceramic fabrics. All the materials showed significant drops in breaking and burst strengths after heat cleaning. Measurable drops in strength did not occur when the fabrics were heat treated for 100 hours at 1,650°F or for 500 hours at 1,650°F. The large drop in strength from as-received to heat-cleaned samples indicate that the organic components in the ceramic fabric may have a significant effect on the ultimate strength of the fabric. Since no organic components will remain during applications at high temperature, all strength measurements for ceramic fabrics should be conducted on heat-cleaned material only. The tensile test and Mullen Burst are actually similar tests since they both stretch the fabric to the point of failure. In fact, the Mullen Burst values for a fabric can be derived from the tensile break strength and elongation measurements.

MIT FLEX FOLD TESTING

Because ceramic fabrics are more brittle than conventional fabrics, the folding angle and tensioning load were decreased below those used for the conventional MIT fold endurance test. A Model No. 2 tester with a dead weight used for tensioning the specimen compensates for any effects due to specimen elongation. The nominal fold angle used was 45° in both directions, as compared to 135°, which is specified for testing conventional fabrics. Results for tensioning loads of 1 to 2 lbs on two types of woven fabrics are presented in Figure 4. A tensioning load of 2 lbs was used in the first set of tests, but this load was reduced to 1 lb to obtain a greater number of flex cycles with the brittle ceramics. Heat-cleaned WN had higher flex endurance compared to heat-cleaned WR, but the flex endurance of WN dropped significantly after heat treatment at 1,650°F. Figure 4 shows that with a 2-lb load, there were no differences in the number of flex cycles before failure between the heat treated WN and WR. With a 1-lb load, WN flex life was significantly greater than the WR material. Thus, a lower tension weight is more desirable for MIT flex testing of ceramic fabrics since it allows higher flex cycles and differences in fabric flex endurance may be detected. It is interesting to note that while the MIT test showed a large decrease in the flex resistance of WN when the fabric was heated from 1000°F (heat cleaning) to 1650°F, the tensile test showed the tensile break strength to remain fairly constant between these two temperatures. This indicates that one mechanical test is insufficient to indicate potential changes with temperature which can occur in a ceramic fabric.

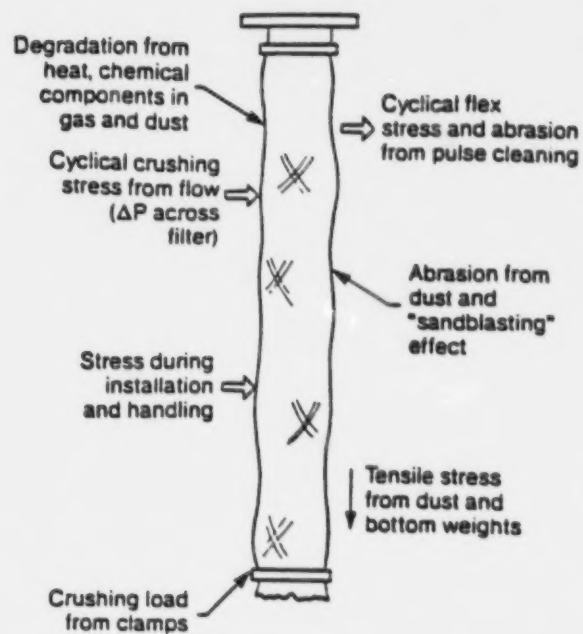


Figure 1. Stress encountered by fabric filter.

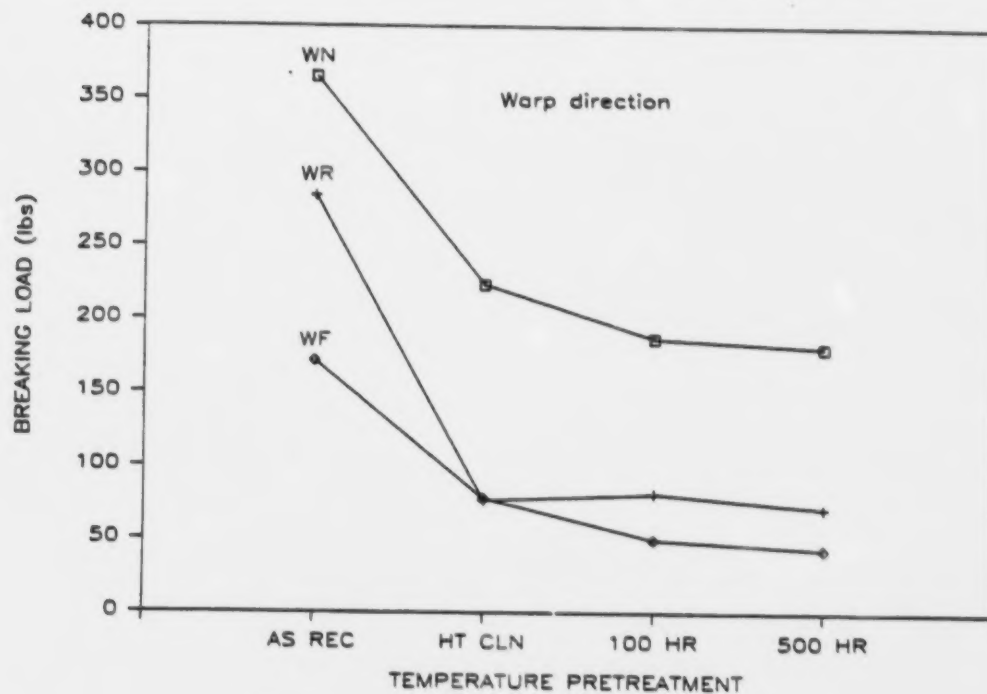


Figure 2. Breaking load tensile test.

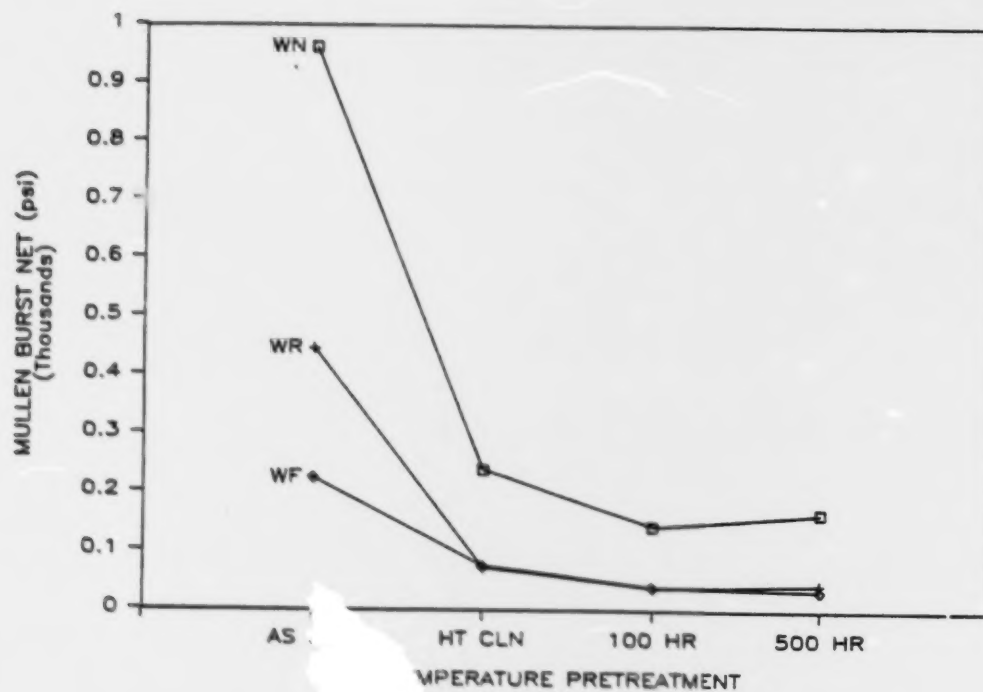


Figure 3. Mullen Burst test.

(45 deg fold angle, warp dir)

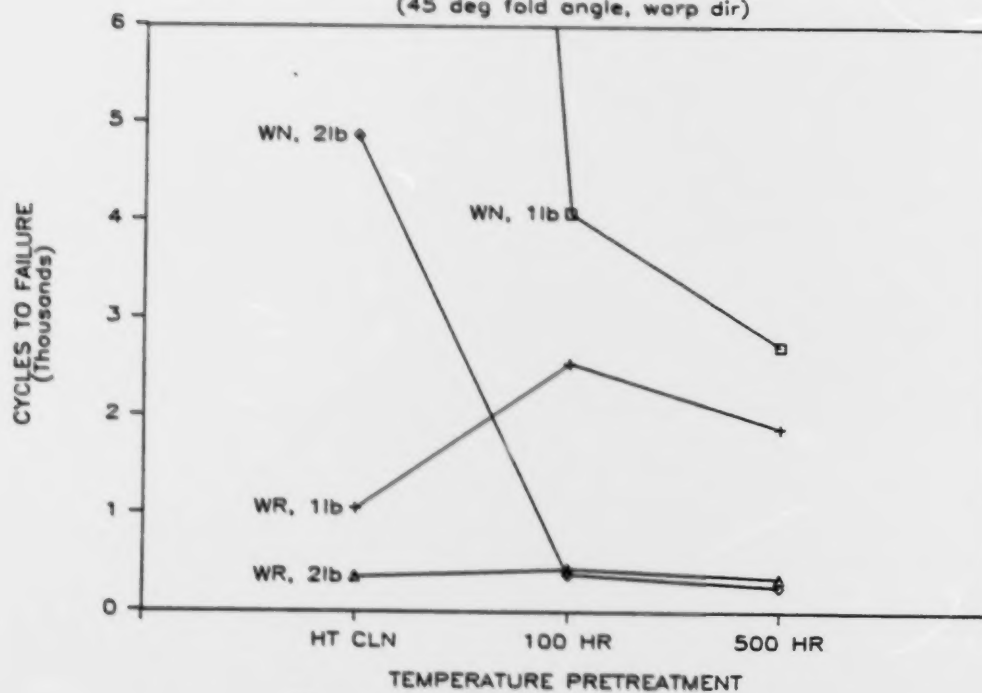


Figure 4. MIT folding test.

CRUSH TESTING

The crush tester is devised to detect changes in brittleness of different ceramic fabrics. Results for six fabrics are summarized in Table 3. The crush resistance of all the ceramic fabrics decreased as a function of temperature and length of exposure at 1,650°F. In the woven fabrics, results are similar to the MIT flex tests where values changed dramatically from heat cleaned to heat treated and remained fairly constant at 1,650°F. WF contains metallic wires as the yarn insert which hold the fabric together even when the ceramic yarns are totally broken. The measured crush resistance values for WF are therefore not accurate. In general, the crush tester provided data with coefficients of variation that range from 17 to 50 percent. This variation is quite good considering that the method is totally new, and only a limited number of samples have been tested.

To characterize the test method further, the crush resistance of ceramic fabrics was tested as a function of the tension weight, crusher weight, and crusher velocity. The crush resistance varied linearly with both the tension weight and the crusher weight. The crush resistance is much more sensitive to crusher velocity where the crush resistance varied as the 3.4 power of velocity. In all cases, however, there seemed to be suitable adjustments available to provide a quick and reliable test for ceramic fabric strength in terms of impact resistance.

HEAT FLEX TEST

The heat flex test is designed as an alternative to the MIT fold endurance test with the added capability of testing fabric specimens at high temperatures. The radius of curvature produced by the heat flex rod is much less severe compared to the curvature of the clamps used in the MIT flex, and the stresses involved during flex testing are easier to analyze for the heat flex test. Results of flex resistance at 1,650°F are presented in Figure 5 for different materials. As in the other cyclical tests, WN has higher values compared to WR. Although WF contained metal wire inserts, it did not seem to have very high flex resistance. To characterize the heat flex tester further, the flex rod diameter and tension weights were changed. In Figure 6, cycles to failure for heat-cleaned WN tested at room temperature is plotted as a function of radius of flex rod/stress produced by tensioning weight (R/S).

Some projections can be made on potential filter life by considering the potential stress levels which may be encountered by a filter fabric during field use and the cycles to failure measured as a function of different stress levels in the heat flex test. For a pulse-cleaned WN filter, assuming that there is a small fold in the filter (0.2-in. slack between filter and support case), the projected life is a minimum of 3,000 pulses at 1,650°F. A filter with less slack or lower cleaning energies will have considerably higher flex life. However, fabric shrinkage at temperature must be taken into account for considering filter slackness requirements. Table 4 shows the shrinkages measured for WN and WR at 1,650°F and the slackness between filter and cage needed for filter shrinkage and cage expansion.

TABLE 1. FABRIC SPECIFICATION AND PERFORMANCE EVALUATION

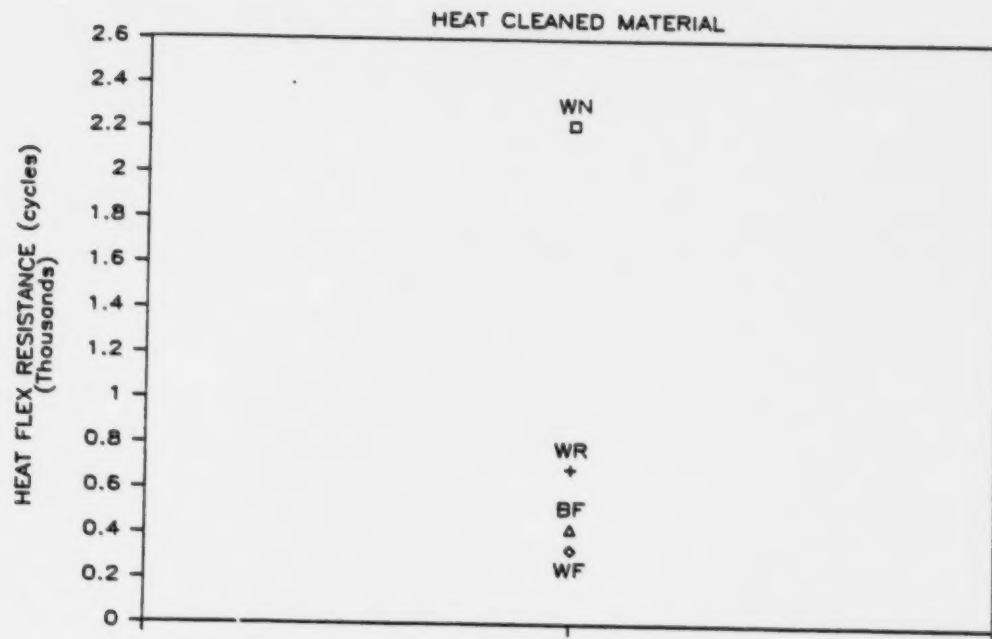
-
- Type of material
 - Weight, thickness
 - Weave, yarn count, warp/fill yarn type
 - Compression and recovery
 - Shrinkage
 - Air permeability
 - Tensile break strength and elongation at break
 - Mullen Burst strength
 - MIT flex
 - Heat flex
 - Crush resistance
 - Collection efficiency and cleanability
-

TABLE 2. PROPERTIES OF FABRIC SELECTED FOR TESTING

Material	Composition	Fiber diameter (μm)	Fabric type
BS	Al_2O_3 , SiO_2	2	Mat
BF	Al_2O_3 , SiO_2	2 to 3	Needled mat
BC	SiC	10 to 15	Mat
WN	Al_2O_3 , SiO_2 , B_2O_3	11	5-harness weave
WR	SiO_2	8 to 10	8-harness weave
WF	Al_2O_3 , SiO_2	2 to 3	2 x 2 twill

TABLE 3. CRUSH RESISTANCE OF CERAMIC FABRICS

Fabric condition	Crush resistance (cycles to failure)					
	WN	WF	WR	BC	BF	BS
As received	>5000	>5000	>5000	1417	1798	9
Heat cleaned	807	>5000	138	0	1973	10
100-hour heat treat	208	2114	179	--	716	1
500-hour heat treat	102	1920	74	--	52	1



TESTED AT 1650 DEG F

Figure 5. Heat-flex test.

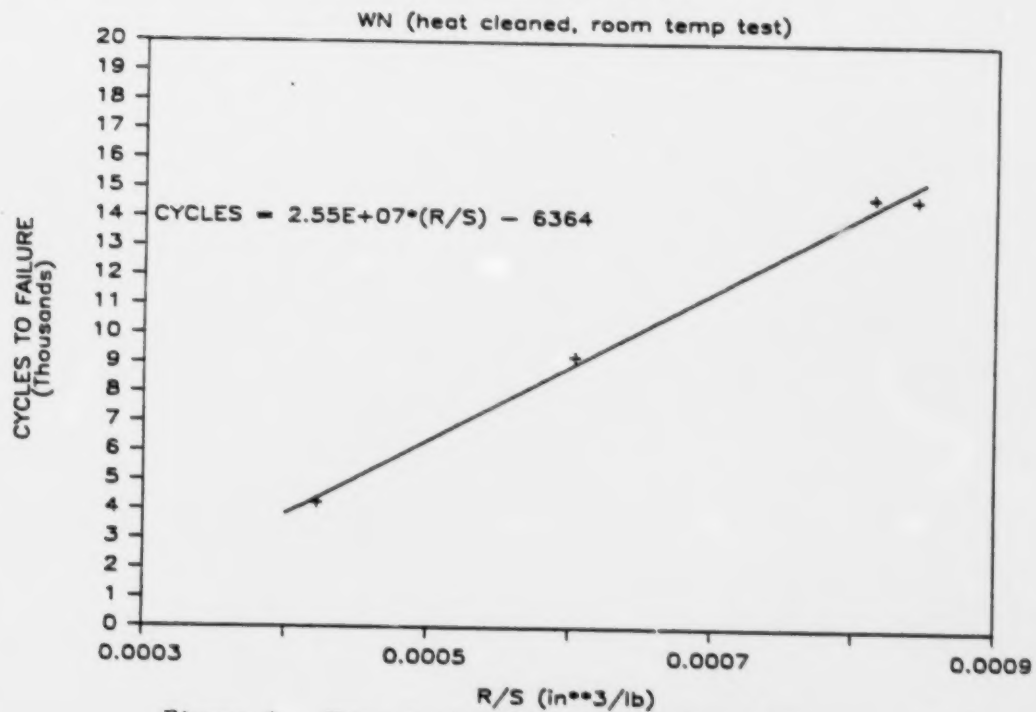


Figure 6. Characterization of heat-flex test.

BENCH-SCALE FILTER TESTS

The bench-scale filter test cell was used to evaluate fabric collection efficiency, cleanability, and short-term durability. The test cell can accommodate filters up to 4 in. in diameter and 1-ft long. Filters can be tested at temperatures up to 1,650°F. Gas flow, dust loading before and after the filter, and filter pressure drop are measured over the desired number of filtering and cleaning cycles.

Typical test operating conditions and results for two woven ceramic fabrics are summarized in Table 5. Dust from a pressurized fluidized-bed combustor was used as the test dust. Collection efficiencies in excess of 99.9 percent were measured consistently for both fabrics.

EVALUATION OF CERAMIC FILTERS FOR GASIFIER APPLICATIONS

An additional program supported by the U.S. DOE Morgantown Energy Technology Center is to evaluate ceramic fabrics for particulate removal in fuel gas from a fluidized-bed coal gasification process. In preparation for tests of bag filters in an actual gasifier gas stream, several candidate filter media were prequalified by determining mechanical properties of ceramic fabrics before and after exposure to a simulated gasifier stream and by evaluating filter performance using the bench-scale test cell.

Based on past experience at Acurex in filtering particles similar to gasifier char, two ceramic fabrics were chosen as candidate filters. These filter candidates were as follows:

- Primary Candidate -- Fabric F, woven ceramic fabric containing composite yarns formed by twisting continuous filament yarns with staple yarns. The resultant fabric has a texturized appearance.
- Alternate Candidate -- Fabric S, a smooth fabric woven from continuous filament yarns in a five-harness weave. Both a single- and double-layer Fabric S were tested.

The candidate filter fabrics were characterized to determine the fabric's weight, thickness, thread count, and yarn and fiber structure. The fabrics were then prepared by exposing them to the following conditions:

- Heat cleaning at 1,000°F for 24 hours
- Heat treating at 1,600°F for 100 hours in air
- Heat treating at 1,600°F for 100 hours in a simulated gasifier environment with gasifier char applied to the fabric
- Heat treating at 1,600°F for 100 hours in a simulated gasifier environment without char applied to the fabric

The simulated gasifier environment was comprised of 13.5 percent carbon monoxide, 10 percent carbon dioxide, 3.5 percent methane, 38 percent

TABLE 4. MATERIAL SHRINKAGE CONSIDERATIONS FOR FABRICS

Material	Linear shrinkage (percent)	Cage expansion (percent)	Slack ^a (in.)
WN	0.5	2	0.05
WR	10.5	2	0.25

^aRequired between filter and cage.

TABLE 5. BENCH-SCALE FILTER TEST OPERATING CONDITIONS AND RESULTS FOR FABRICS WF AND WN

Test parameter/result	Fabric WF	Fabric WN
Filter area (ft ²)	0.78	0.78
Average temperature (°F)	1654	1642
Average pressure (psig)	8.2	8.4
Average A/C ratio (ft/min)	5.2	5.1
Average dust load (gr/acf)	4.0	4.1
Average cleaning efficiency (%)	99.99	99.98
Average ΔP before cleaning (in. H ₂ O)	4.7	9.0
Average ΔP after cleaning (in. H ₂ O)	1.0	3.9
ΔP at start of test (in. H ₂ O)	0.5	2.25
Time to increase to 5 in. H ₂ O (hrs)	3.0	1.8
Cleaning pulse width (sec)	0.05	0.05
Cleaning pulse pressure (psig)	95	95
Cleaning pulse frequency (number/hr)	1	1

nitrogen, 17 percent hydrogen, and 0.3 percent hydrogen sulfide with trace amounts of alkali vapor.

After the fabrics were exposed to the preceding conditions, fabric samples were tested for flex resistance at room temperature, crush resistance, and tensile strength. Results of these tests are summarized below. It should be noted that since the flex tests and the tensile strength tests showed the same general trend, only the flex results are presented.

CRUSH-RESISTANCE RESULTS

Table 6 shows the results for the crush-resistance measurements. Both materials showed a drop in crush resistance after being heated up to 1,600°F. The crush resistance appeared to depend mainly on the temperature pretreatment of the ceramic materials. There was no difference between materials treated at 1,600°F in air and materials heated at 1,600°F in the gasifier simulator in the reducing environment.

FLEX-RESISTANCE RESULTS

Table 7 shows the values measured for the flex resistance. Fabric S showed a flex resistance much higher than that of Fabric F, even though F contains some of the same fibers that were used to weave Fabric S. Both fabrics showed a drop in flex resistance with temperature pretreatment. For Fabric S, char appeared to give a positive effect as if it were acting as a lubricant during the flexing process. No lubricating effect of char was observed for Fabric F.

Without char, the flex resistance of Fabric S appeared to be independent of the gaseous environment of exposure. These results, together with crush-resistance measurements, seemed to indicate that the gasifier environment should not have a detrimental effect on the two fabrics. In fact, in some cases the presence of char may actually improve flex life of the filter fabric.

The filtration performance of different filter fabrics for gasifier char was evaluated in a bench-scale filter facility. As a baseline, the performance of these fabrics using PFBC dust was also measured.

TEST DUST DESCRIPTION

The two dusts used to test the potential filter medium at 1,600°F were flyash produced from a pressurized fluidized-bed coal combustor (PFBC) and char produced from a coal gasifier. Their density and size distribution are described in Table 8. The PFBC dust has a higher bulk density and particle density. During the bench-scale test, both dusts were fed at the same mass feedrate with an inlet dust loading of about 4.5 gr/acf.

TABLE 6. RESULTS OF CRUSH-RESISTANCE TESTS

Material	Crush resistance (cycles)	Standard deviation	Coefficient of variation	Number of samples
S/ht clean	1079	610	0.56	6
F/ht clean	1367	264	0.19	5
S/air ht	565	302	0.54	6
F/air ht	435	131	0.30	5
S/sim w/ch	569	142	0.25	6
F/sim w/ch	435	124	0.29	5

Notes: Fabric S is a single-layer weave.

ht cln -- heat cleaned.

air ht -- kept at 1600°F in air for 100 hours.

sim w/ch -- in gasifier simulation can for 100 hours at 1600°F
with gasifier char spread on the fabric surface.

TABLE 7. RESULTS OF FLEX-RESISTANCE TESTS

Material	Flex resistance (cycles)	Standard deviation	Coefficient of variation	Number of samples
S/ht clean	13,807	2,930	0.21	6
F/ht clean	1,924	66	0.03	4
S/air ht	5,622	2,330	0.41	6
F/air ht	441	60	0.14	5
S/sim w/ch	10,925	3,614	0.33	6
F/sim w/ch	450	52	0.12	5
S/sim w/o ch	6,044	2,509	0.42	6

Notes: Fabric S is a single-layer weave.

ht cln -- heat cleaned.

air ht -- kept at 1600°F in air for 100 hours.

sim w/ch -- in gasifier simulation can for 100 hours at
1600°F with gasifier char spread on the fabric
surface.

sim w/o ch -- in gasifier simulation can for 100 hours at
1600°F without gasifier char spread on the
fabric surface.

TABLE 8. PHYSICAL PROPERTIES OF TEST DUSTS
 PARTICLE SIZE DISTRIBUTION AS MEASURED BY
 COULTER COUNTER TECHNIQUE

Particle size distribution (microns)	Mass percent greater than given size	
	PFBC dust	Gasifier char
0.35	99.90	
0.54	99.00	
1.00		100.00
1.08	94.00	
1.26		98.19
1.59		95.49
2.00		91.56
2.25	78.00	
2.52		85.78
3.17		77.73
4.00		67.73
4.50	50.00	
5.04		55.96
6.35		43.31
8.00		27.71
8.02	22.00	
10.08		13.24
12.70		3.48
13.23	6.00	
16.00		0.49
20.20		0.00
23.57	1.00	
32.08	0.10	
Bulk density	56.1 lb/ft ³	31.0 lb/ft ³
Particle density	2.87 gm/cc	2.18 gm/cc

TEST OF FILTERS ON PFBC DUST

Filter materials S and F were first tested using PFBC dust because of prior experience filtering PFBC dust with fabrics similar to these. For both cases, the dust removal efficiency was found to be greater than 99.99 percent. The pressure drop across the filter after cleaning was constant (i.e., there was no evidence of filter blinding over the 50-hour test period). At the end of 50 hours of testing, no excess weakening of the bags was noticed, and the dust cake was evenly coated over the bag. However, the dust cake on Fabric S was not as thick as the cake on Fabric F.

TEST OF FILTERS FOR GASIFIER CHAR

Using gasifier char, the dust removal efficiency for Fabric F showed some variation (99.65 to 99.96 percent) for the first 30 hours of the test, but eventually stabilized at 99.99 percent for the remaining 45 hours. During the test the filter was cleaned once every 0.5 hours rather than once per hour as when filtering PFBC dust. The pressure drops across fabric were about 4.0 and 0.6 in. of water before and after cleaning, respectively, at the beginning of the 75-hour test period, and rose to 5.8 and 1.8 in. of water, respectively, by the end of the test. These data indicate a significant increase in pressure drop over a short period, but the pressure drop did appear to be stabilizing. Longer term tests are required to establish a stable operating point.

While removing the cage assembly from the test cell, the char fell off the filter bag much easier than it had for PFBC dust. The char does not appear to form as good a cake as does PFBC dust. The dust layer was thinner than had been found when using PFBC dust. There appeared to be some pin holes in the filter cake. From visual inspection of the inside of the bag, there was no significant leakage through the filter.

The smooth S fabrics did not perform as well as Fabric F for collecting gasifier char. Both single-layer Fabric S and double-layer Fabric S were tested. For PFBC dust, both fabrics had shown excellent dust removal efficiency. However, the gasifier char was significantly more penetrating even though the effective mass media diameters of the two dusts were similar. After startup, the dust removal efficiency for double-layer S fluctuated between 89 and 99 percent. After 20 hours of operation, a value of 97.4 percent was measured. During its first 8 hours of operation, this fabric also showed some sharp increase in pressure drops, but the pressure-drop appeared to stabilize after the first 8 hours. Based on past experience when filtering PFBC dust, these pressure-drop fluctuations along with the relatively low collection efficiency seemed to indicate the possibility of a broken filter. However, post examination of the filter showed no apparent damage and no leakage around the clamp regions. A significant number of pin holes were observed. From visual inspection of the back side of the filter, there appeared to have been some dust penetration. Single-layer Fabric S had a low dust removal efficiency fluctuating from 12 to 64 percent. The efficiency did not seem to improve while the pressure

drop across the filter continued to increase. The test was discontinued after 8 hours.

Although Fabric S exhibited superior mechanical properties compared to Fabric F, Fabric F was selected for the gasifier tests because of its ability to collect the char. It may be possible to obtain adequate collection efficiency with Fabric S by modifying the weave and fiber structure, but this investigation was beyond the scope of the present study.

CONCLUSIONS

The small-scale testing of ceramic fabrics has proved very useful in characterizing potential fabrics for high-temperature particle control for PFBC and gasifier applications. General conclusions drawn from the tests are as follows:

- Conventional test methods for textile fabrics were readily adaptable to the evaluation of ceramic fabric with only minor modifications
- Three new test methods developed for ceramic fabrics provided useful information for characterizing ceramic fabrics. These tests were the crush-resistance test, the heat flex test, and the bench-scale filtration test. While the heat flex and bench-scale filtration tests are analogous to conventional fabric tests, these two tests are different in that they can assess the affect of the high temperature environment on ceramic fabrics.
- As-received ceramic fabrics contain organic sizing that can mask the actual properties of the fabric at room temperature: all fabrics should, therefore, be heat cleaned before testing
- The mechanical properties of ceramic fabrics are temperature-dependent. The maximum operating temperature will be a key parameter determining ultimate filter durability.
- The mechanical properties of ceramic fabrics are also time-dependent when exposed to high temperature. The changes in mechanical properties with time are gradual, which may have an effect on filter durability in the long term. Longer term tests are required to assess the effect of temperature and environment.
- All ceramic fabrics tested exhibited excellent collection efficiency for PFBC dust
- Gasifier char is significantly more penetrating than PFBC dust and has very low cohesivity. It is apparent that fabrics are likely to perform differently in the two diverse operating environments of PFBC and gasification.

- The gasifier environment alone did not have a significant effect on the mechanical properties of the filter fabrics tested
- It is important to select the proper fabric weave and fiber structure for a specific application. Screening tests clearly indicate that Fabric S was unacceptable for the collection of gasifier char and that Fabric F was an acceptable filter because of its high collection efficiency.
- Gasifier char may have a lubricating effect for some fabrics, thus extending the flex life of the fabrics

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HOT GAS CLEAN-UP BY MEANS OF POROUS CERAMIC FILTER ELEMENTS

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ABSTRACT

This technical paper addresses applications whereby the use of ceramic filter elements is suitable in the removal of particulates from hot gas streams. Particle sizes typically encountered range from 0.2 to 15 microns, and a wide array of equipment types have been proposed or used in particulate removal, depending upon process conditions. Materials commonly used in the construction of ceramic elements are listed, along with inherent advantages and disadvantages of each. Shapes and dimensions of typical elements are discussed, as are currently available grades and pore sizes.

The relationship between pressure drop and gas velocity associated with the use of ceramic filter elements changes with the length of time the elements remain in use. The incremental change in pressure drop between subsequent filter cycles of a ceramic element approaches zero as the total number of cycles increases, provided gas velocity remains constant. Eventually, the filter element reaches an "equilibrium" state, in which the pressure drop vs gas velocity relationship remains unchanged.

Pilot test work performed at the Institute of Gas Technology's U-Gas Plant in Chicago, Illinois is discussed, in which a single silicon carbide filter element was used to remove particulates from a coal gasification product stream. The anticipated correlation between data taken under process conditions vs data generated under ambient conditions proved to be highly accurate.

Multiple element filter configurations are addressed in the paper, as is the typical "jet-pulse" cleaning method of the ceramic elements.

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Highly efficient particulate removal from hot gas streams is important in a number of process points including:

- A. Preparation of combustion "off" gasses for re-use across secondary turbines.
- B. Removal of particulates from hydrocarbon processing gas streams above the dew point.
- C. Removal and/or recovery of catalysts from "off" gasses.
- D. High quality cleaning of emissions from "radwaste" incinerators.

A lot of "cheap work" can be done on particulate-laden gas streams before the use of porous ceramics. Two cyclones in series will probably remove 70-85% or more of the dust load.

Typically particle size distribution of the remaining dust will be 0.2-15 micron⁽¹⁾ ⁽²⁾ The most commonly occurring particle will be about 4 microns, as shown in Figure #1.

Cyclones aren't the only suitable method of particulate removal. Nor are porous ceramics. Other approaches (3) (4) (5) (6) (7) (8) (9) (10) (11) referenced have varying temperature capability, as indicated in Figure #2. This figure also shows that porous ceramics are well suited to operation above (say) 1600°F. (871°C.). Porous ceramics have existed for a goodly period of time. Still the application of porous ceramics to hot gas filtration has not been widely practised. The purpose of this paper is to tell more about porous ceramics and their effective use.

A. Shapes: In the field of gas filtration, porous ceramic materials are generally available in three shapes, as shown in Figure #3. A. Plates, B. Cylinders, C. Candles (flanged).

Plates (or discs) are most commonly used as gas diffusers, filter bottoms, or bed bottoms. Generally, they are 1-3 centimeters (.4-1.2 inches) in thickness, and some type of support, such as metal, is required.

Cylindrical filter elements are generally 60-70 millimeters (2.4 - 2.8 inches) in diameter, and have a wall thickness of 10-15 millimeters (.4-.6 inches). Despite their attractive cost in comparison to flanged filter candles, cylinders tend to be difficult to work with. Metallic draw rods are required for proper fixing of the cylinders within the filter. Doing so introduces problems associated with differences in coefficient of thermal expansion, which may ultimately lead to premature equipment failure.

Flanged filter candles offer a unique alternative to cylinders or plates. The candles, which are also cylindrical in shape, can be manufactured with an impervious section (or plug) on one end, and a flanged opposite end. The flange end allows for mounting of the candles without the need for draw rods, etc. Both the "plugged" end and the flanged end of the candle are of ceramic composition, and are "fired" in place during candle manufacture.

TEMPERATURE

ASH PRODUCED AT 900 - 950 C (1650 - 1750 F)

GAS STREAM CONTAMINATION LEVEL

PFBC DUST FROM COMBUSTOR AT 7,000 - 20,000 ppm
CE/WIS DUST FROM COMBUSTOR AT 1000 ppm

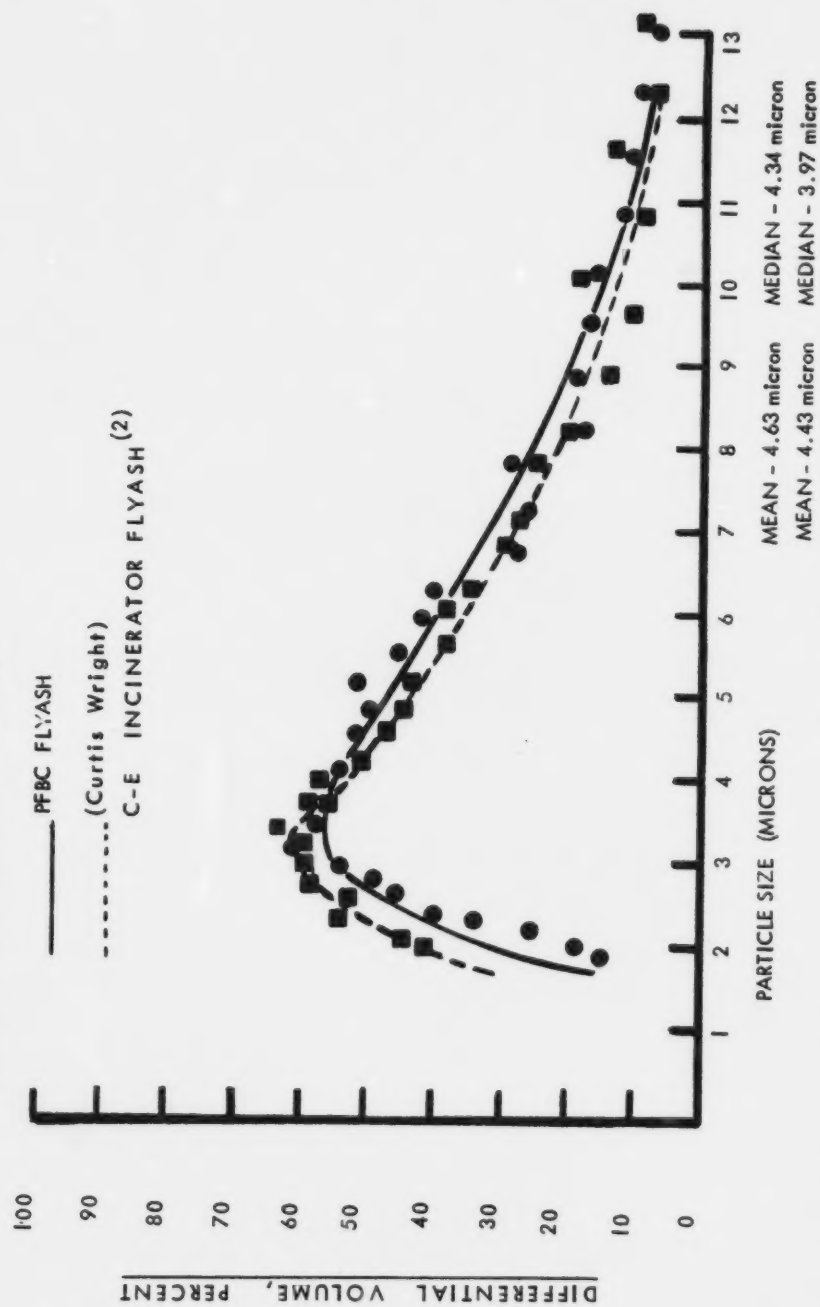


FIGURE #1 - ASH COMPARISON

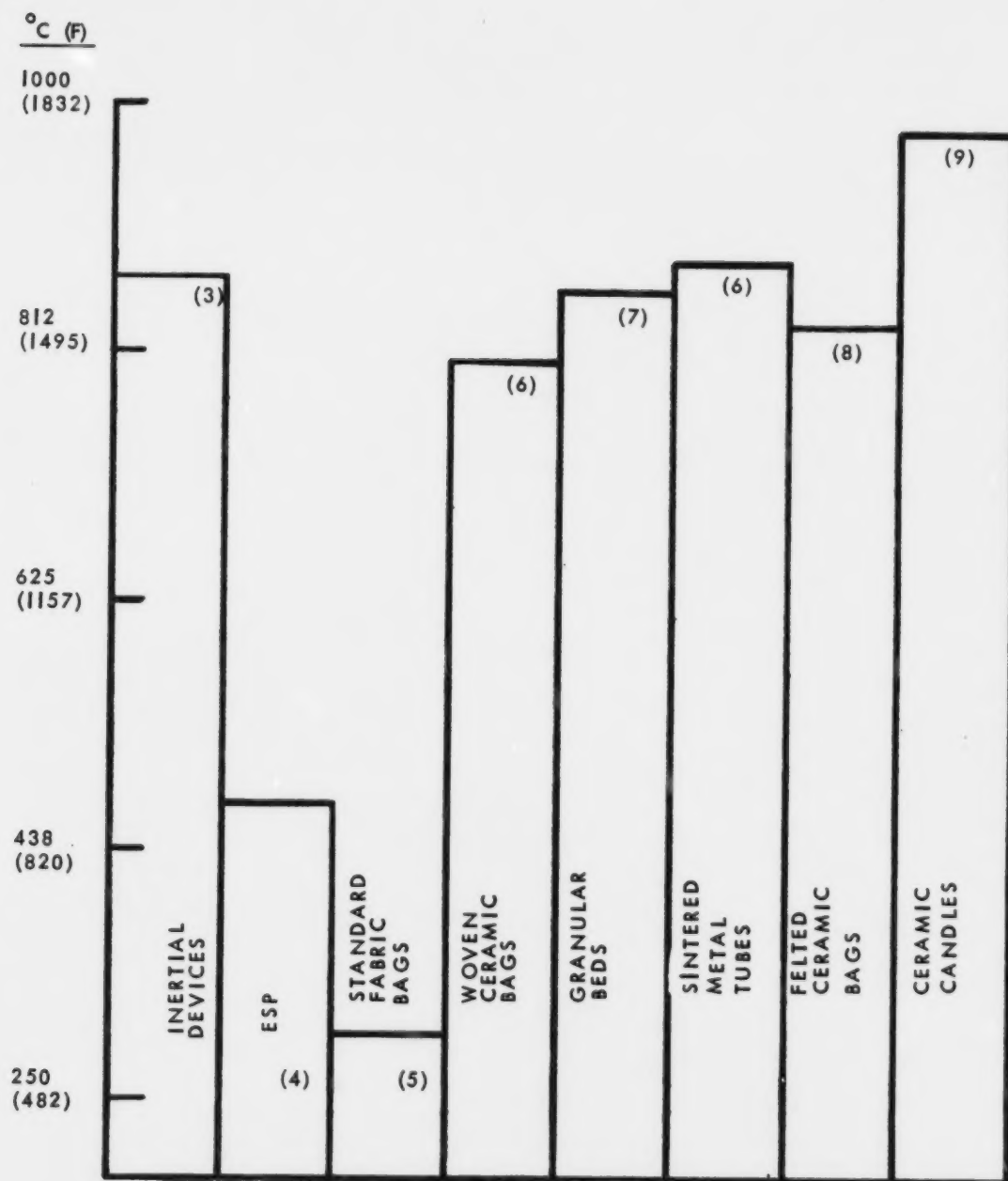


FIGURE #2 - TEMPERATURE CAPABILITY

POROUS CERAMIC SHAPES

1. PLATES - DIFFUSERS, FILTER BOTTOMS USUALLY
1 - 3 Cm THICK
COMMONLY NEED METAL SUPPORT
2. CYLINDERS - FILTER ELEMENTS
USUALLY 60 - 70mm DIAMETER
WALL THICKNESS 10 - 15mm
NEED METAL DRAW RODS
3. CANDLES - FLANGED MONOLITHIC STRUCTURE
DO NOT NEED DRAW RODS

FIGURE #3

The resulting product is a strong filter element , virtually free of internal stresses or flaws. For lengths longer than 1 meter (39+inches) such candles can be monolithic or "jointed" with equal quality.

Filter candles are generally mounted within a filter chamber by suspending the candles from a tube sheet. The flanged end of the candle can be recessed within the tube sheet opening, and sealed in place with a high-temperature cement. The cement is cured in place during start-up of the equipment, preventing both the movement of the candle during flow transitions and the migration of particulates past the flanged surface and into the effluent stream. Alternately, high temperature gasketing and a hold down system may be employed.

In practice, filter candle lengths of up to 1.5 meters (59 inches) have been secured in this manner, with highly successful results. As mentioned earlier, wall thicknesses of 10 and 15 millimeters (.4-.6 inches) are typical. The thinner wall thicknesses yields a lower pressure differential at a given gas velocity.

B. Materials: The most commonly available materials of construction for porous ceramics generally fall into one of the four following categories: A. Silicon carbide-SiC, B. Aluminum oxide- Al_2O_3 , C. Quartz-SiO₂, D. Mullite- $3\text{Al}_2\text{O}_3 \cdot 2\text{SiO}_2$. Each of these materials produces ceramic elements with quite different characteristics. For example, an element formed from silicon carbide weighs almost twice that of a like-sized quartz element - a factor that can be quite important when structural support is considered.

However, a silicon carbide filter element is stronger than either a quartz, aluminum oxide, or Mullite element. Quartz is not feasible when the operating temperature exceeds about 400°C. (750°F.). The other three candidates are suitable to to about 1000°C. (1830°F.).

Silicon carbide elements possess excellent thermal shock resistance, but are also the most expensive. The relative costs for the materials are as follows: Quartz/Mullite/aluminum oxide/silicon carbide are approximately 1.0/1.12/1.3/1.75 in price relation. Figure #4 summarizes these important considerations.

Figure #5 illustrates the wall thickness effect mentioned earlier for Grade 20 silicon carbide elements of 10 millimeters and 15 millimeter (.4 and .6 inches) wall thicknesses. A superficial gas velocity ("face" velocities) of up to 20 meters per hour (1.1 feet per minute) has been used for the illustration. Even at this relatively low velocity, the disparity in pressure differential is significant.

C. Grades/Mean Pore Size: The term "grade" was mentioned earlier in reference to the illustration given in Figure #5. The "grade" of a porous ceramic element is commonly defined as the quantity of air in scfm (1.7) (quantity of air in scmh) that will pass through 1 square foot (930 square centimeter) of the ceramic at a differential pressure of 2 inches of water (0.5 kilopascals). As the "grade" of the element decreases numerically, so does the air flow.

CERAMIC MATERIALS
POROUS CERAMIC
 MATERIALS OF CONSTRUCTION

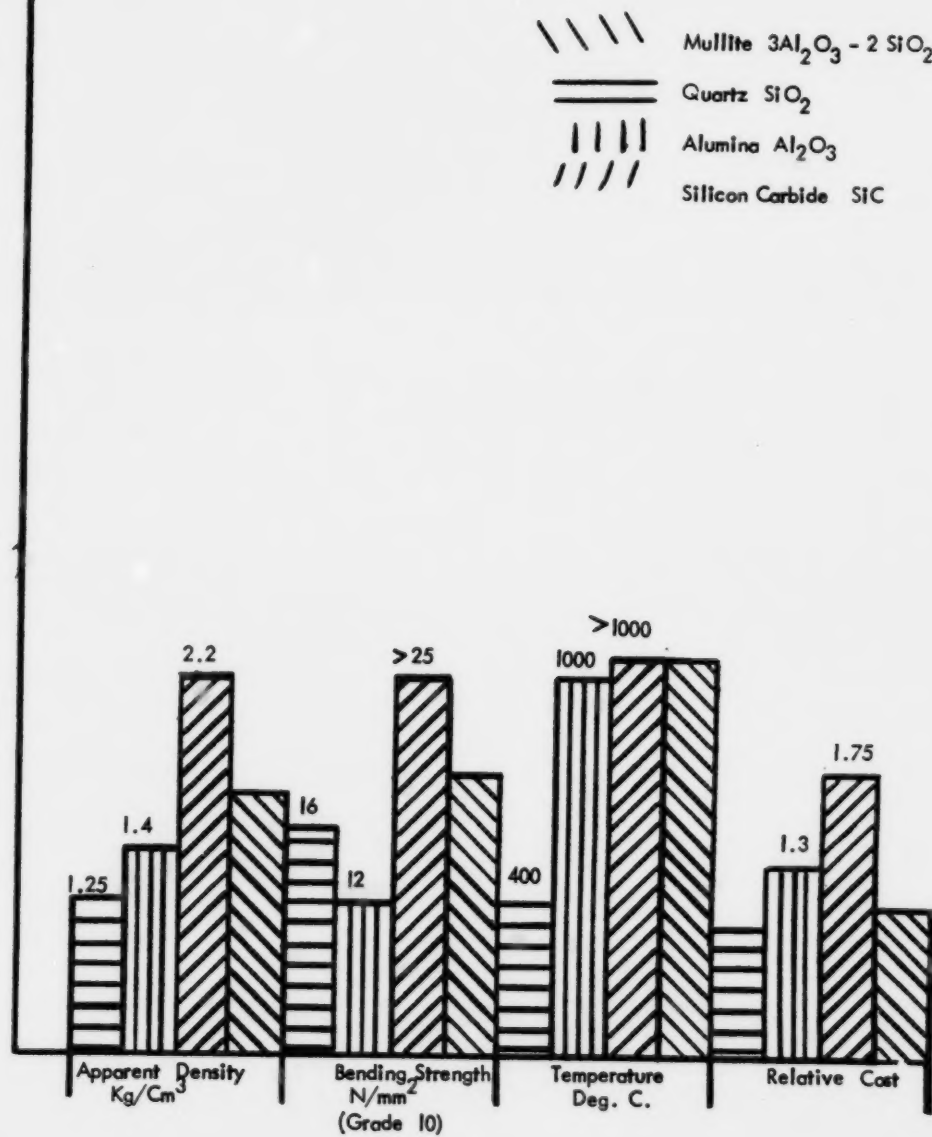


FIGURE #4

POROUS CERAMIC CANDLES
EFFECT OF WALL THICKNESS
(SIC GRADE 20)

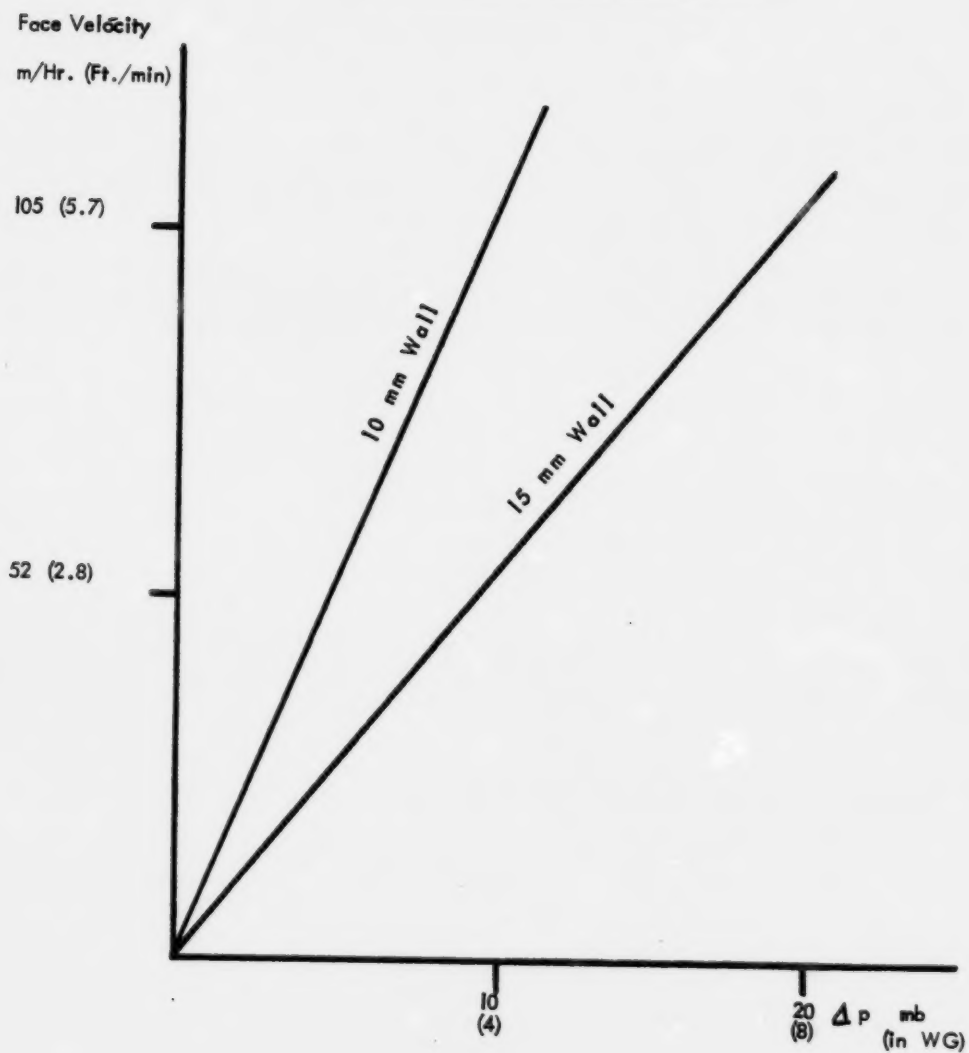


FIGURE #5

Gas filtration applications typically call for grades in the range of Grade 5 to Grade 60. As shown in Figure #6, there is an approximate linear correlation between the grade of a particular element, and mean pore size associated with that element. It must be emphasized; however, that this relationship will vary somewhat with material composition and method of manufacture.

D. Pressure Drop/Face Velocity/Dust Load: Figure #5 showed earlier the effect of wall thickness on the pressure differential across the filter element, at varying face velocities. At any given wall thickness, then, it stands to reason face velocity also has an effect on pressure drop, as does cake thickness.

Typically, any filter candle at any given face velocity will show a particular pressure differential when tested "as manufactured" in the absence of dust. See Figure #7.

After a number of service cycles, during which the filter candle is repeatedly subjected to gas/particulates (in the service flow direction), followed by a high pressure pulse of gas to clean the filter candle (opposite of the service flow direction), the candle reaches an "equilibrium" state with respect to pressure differential and face velocity. The attainment of this equilibrium state is dependent upon the process conditions encountered, filter candle composition, candle grade, particulate characteristics, etc. Pilot studies have shown that this state is, for all practical purposes, achieved after about 200 service cycles.

Prior to reaching the equilibrium state, the pressure differential across the candle in the service flow direction increases at a faster rate over subsequent service cycles (at constant face velocity, particulate loading, etc.). The effect of this transitional state is observed as either an incremental decrease in service cycle lengths (for service cycles terminated by a differential pressure set-point), or by an incremental increase in the maximum pressure differential value (for service cycles terminated by a set time interval).

Once the equilibrium state is achieved, the filter candle exhibits a pressure differential roughly twice that of the "as manufactured" pressure differential, when tested at the same face velocity in the absence of particulates. This is shown in Figure #7. This figure also shows the pressure differential vs face velocity relationship of a filter candle with 2.5 millimeters (0.1 inches) of cake in place. Both the "clean" filter pressure differential relationship (denoted as the equilibrium state), and the maximum pressure differential relationship (denoted as the equilibrium state), and the maximum pressure differential relationship are extremely important in equipment sizing and specification.

Operation of a Single Candle Test: Porous ceramic element selection is best determined by the utilization of a pilot test filter under actual process conditions. A good example of information obtained from actual pilot test

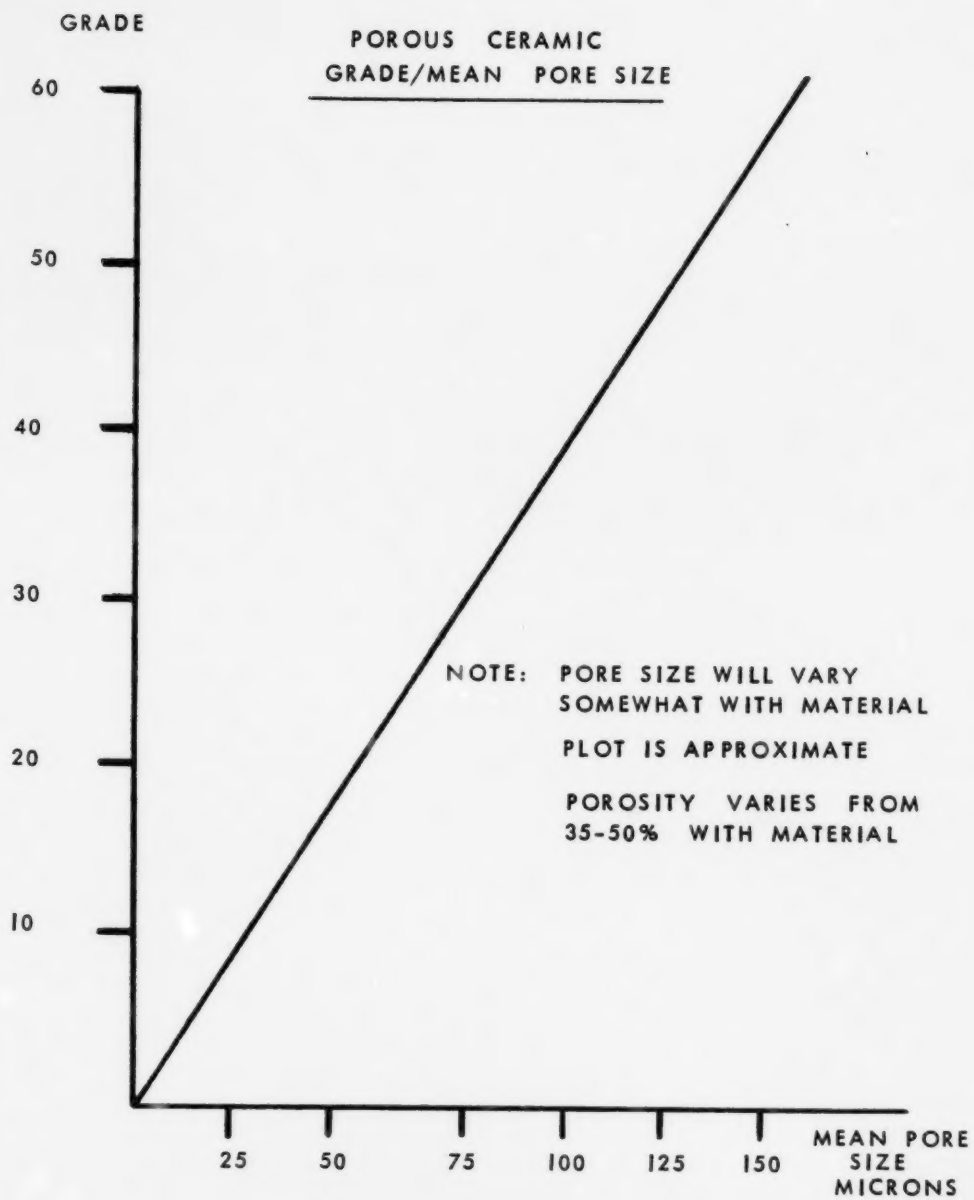


FIGURE #6

POROUS CERAMIC
PRESSURE DROP / FACE VELOCITY
(SiC - 20)

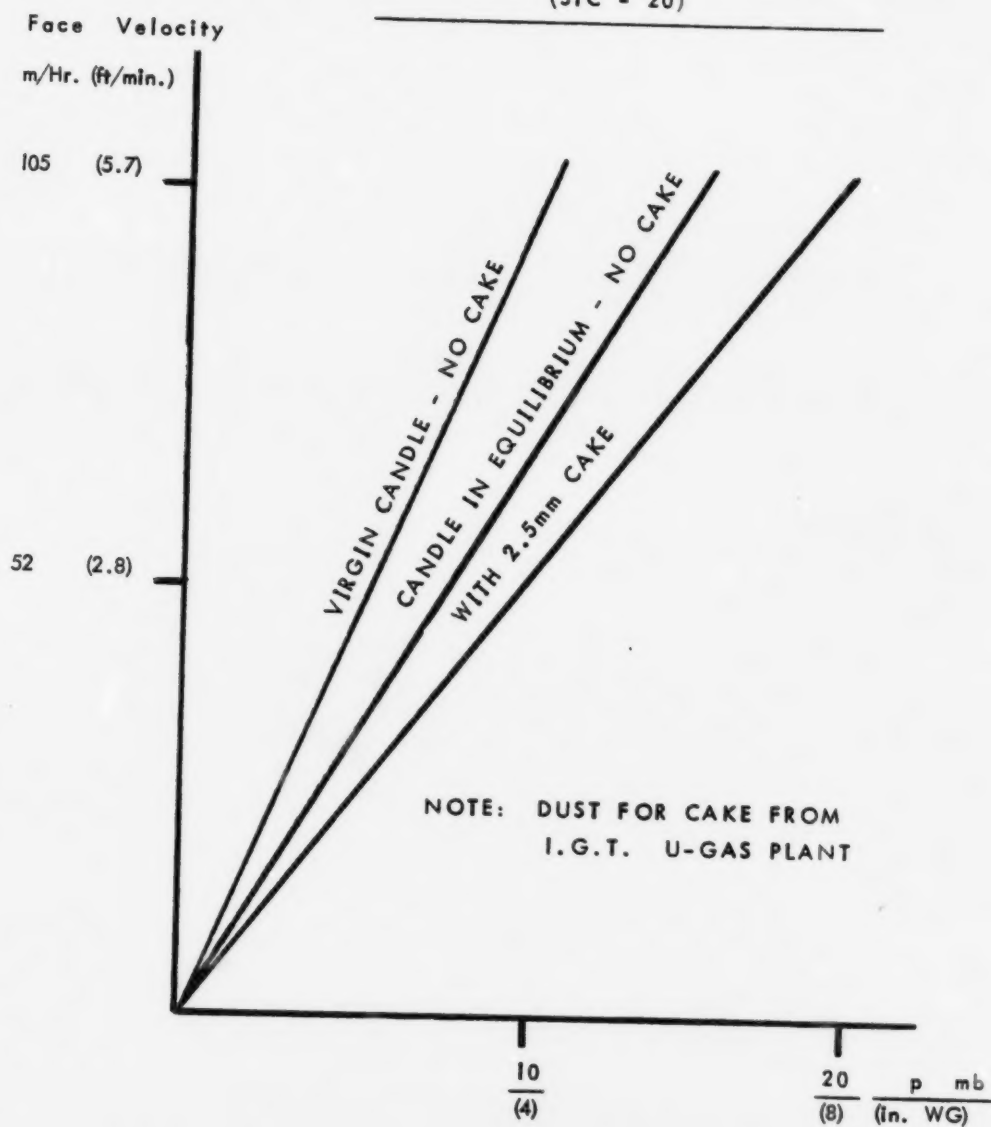


FIGURE #7

data is given in Figure #7, which shows data plots derived from tests performed at the Institute of Gas Technology "U-Gas" Plant in Chicago, Illinois in 1984. (12) (13) (14) (15)

Research has shown that data taken at ambient conditions can be correlated with data taken under actual process conditions by application of equations of state (provided flow rates, particle loading, etc., remain constant). If actual pilot tests under service conditions cannot be made the correlation can be relied upon for sizing. Tests incorporating a "pilot" filter are preferred, particularly where scale-up to a multiple element configuration is envisaged.

The use of a single candle on-line test filter at the Institute of Gas Technology's U-Gas Plant in Chicago, Illinois is illustrated in Figure #8. The test was conducted to confirm that a single candle filter under high temperature conditions would yield results similar to that which could be generated by application of equations of state on filtration data taken under ambient conditions.

The U-Gas Plant equipment sequence consisted of a full-scale coal gasifier vessel, 1st stage fines collector/fines re-cycle, 1st stage quencher, 2nd stage fines collector, 3rd stage fines collector, 2nd stage quencher and Venturi scrubber. A fluidized bed of coal was reacted with steam and air/oxygen to produce a "low btu" product gas to be used as fuel.

The single candle test filter, located just after the 1st stage fines collector, was exposed to a "slipstream" of product gas/fines during a six day test. Periodic back-pulsing of the filter element was accomplished automatically by the utilization of an interval timer.

On-line temperature, flow, pressure and particle sensors were used to monitor process conditions/performance. Inlet pressure to the test filter was between 40-50 psi (275-345 kilopascals). Inlet temperatures varied between 600-1100° F. (315-593° C.).

Throughout the test, the total gas flow to the filter ranged from 3 to 6 lb. moles per hour. Gas velocities ranged from 200-350 meters per hour (11-19 feet per second). Service cycle durations ranged between 2 and 10 minutes, depending upon the particulate load to the filter, with a backpulse duration between service cycles of two seconds. It is important to note that the filter never went off-line even for back pulsing.

Because the test filter had a limited dust holding capacity beneath the filter element, a lock hopper arrangement was employed for the removal and cooling of dust from the filter chamber. See Figure #8. Every hour, the lock hopper was isolated from the filter, vented, and purged of dust before being returned to service. The amount of dust collected in the dust drum was tabulated as was the data collected from the on-line sensors.

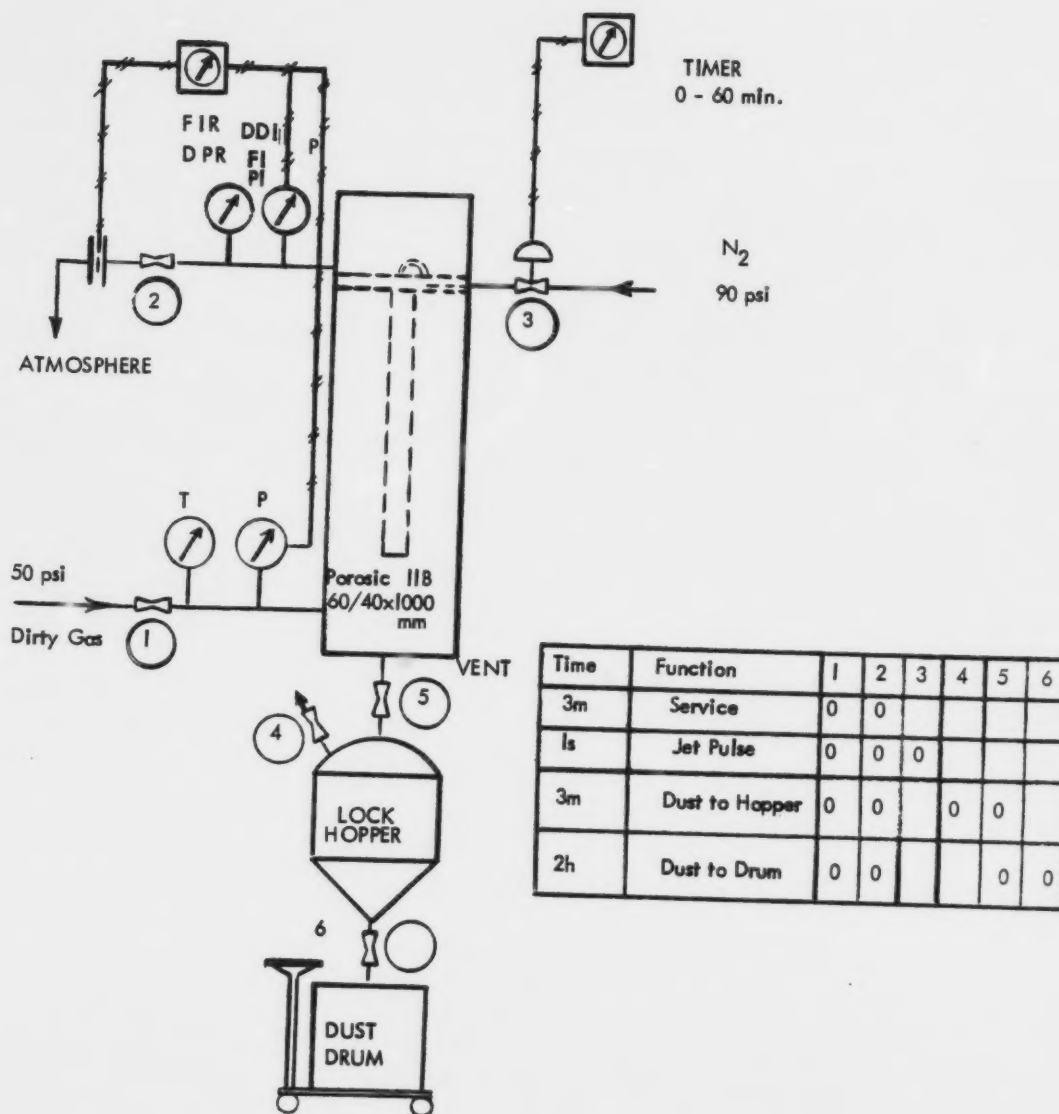


FIGURE #8

Because the filter would be continuously on-line there was concern. There was also some concern about condensation and debris associated with gasifier start-up might "blind the porous element". However, both concerns were dispelled after successful start-up of the gasifier. Throughout the duration of the test, overall operation of the pilot filter was excellent. The filter candle reached an equilibrium state after approximately 150-250 service cycles. A total of over 4000 service cycles were noted, and the amount of particles passing through the filter smaller than 10 micron was undetectable by both the on-line particle sensor as well as a Balston "Microfibre" absolute filter rated at 93% efficiency down to 0.1 micron.

Multi-Candle/Multi-Unit Configurations: Figure #9 depicts back pulse valve sequencing for a multi-candle unit. It differs from valve sequence for a single candle unit. The filter element in the single candle configuration is continually exposed to the influent gas stream, even during back pulsing of the element. This is not the case with multiple element filters, as "unpulsed" filter candles (or rows of candles) tend to absorb the momentary incremental increase in gas flow. In both cases, no difficulty has been experienced at gas velocities of up to 400 meters per hour (21.9 feet/minute). (11)

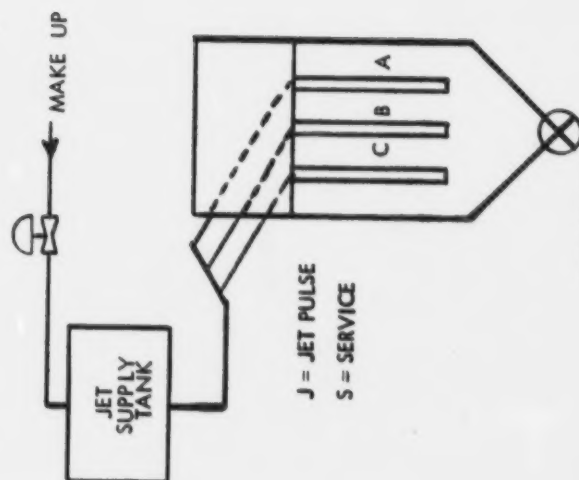
Lastly, Figure #10 illustrates a typical station layout planning sheet for a multiple candle filter system.

Conclusion: A wide array of porous ceramic filter elements are available for different types of gas filtration. Candle material selection is generally a function of temperature/thermal shock resistance requirements. Physical strength and filter candle weight are secondary considerations. Either cylindrically shaped elements or flanged candles can be used, depending on operating temperature. Filter element grade and pore size specifications are dependent on the allowable pressure drop and dust characteristics.

There is a strong correlation between filtration data taken at ambient conditions and data taken under actual process conditions. Equipment sizing can be accomplished in a straightforward, numerical manner, provided that the process parameters required are available for the calculations.

As the use of sophisticated high temperature incineration and combustion technology increases, so will the need for filtration equipment capable of meeting the severe operating conditions. Given their high efficiency and thermal resistance and chemical resistance, porous ceramic filter candles are and will continue to be desirable candidates for such applications.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.



JET PULSE TO GROUPS

OPERATION	Δt	A	B	C	Etc.
CONTINUOUS	$\delta \text{ sec. (MAX.)}$	J	S	S	S
		S	J	S	S
		S	S	J	S
		S	S	S	J
		J	S	S	S

COMMENT:

System-trigger on time $a/o \Delta p$ once systems triggered each group jets sequentially A, B, C, etc. until back to "A" at which time the time Δp trigger is reset.

Minimum sequence time between groups governed by recovery speed on jet supply tank.

For planning purposes estimate 4 cycle/hour/filter.

HGCU FILTERS SERIES 114
JET PULSE SYSTEM

FIGURE #9

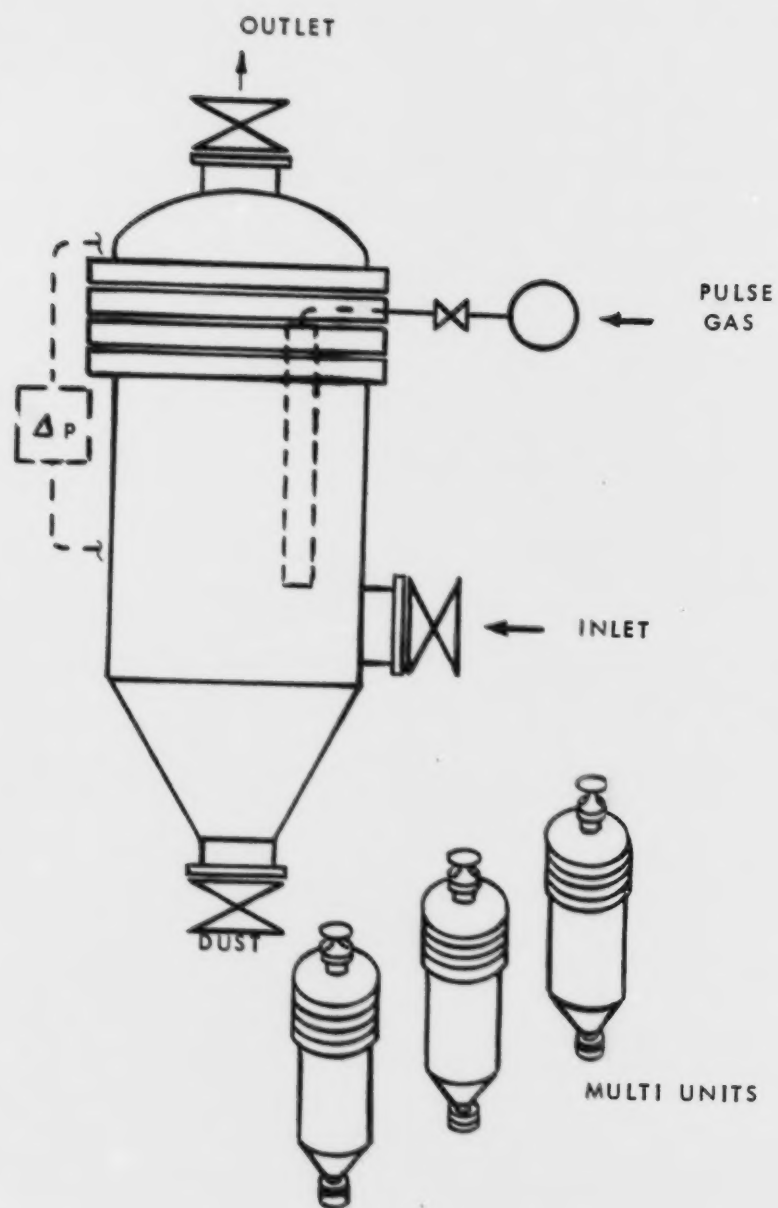


FIGURE #10

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Session C-4: ADVANCED ENERGY APPLICATIONS II

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RESULTS OF PARAMETRIC TESTS ON AN ELECTROSTATIC PRECIPITATOR
OPERATING AT HIGH TEMPERATURE AND HIGH PRESSURE CONDITIONS

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ABSTRACT

The final results of pilot scale tests on an electrostatic precipitator (ESP) operating at pressurized fluidized bed combustor (PFBC) conditions are presented. The tests were performed at the DRI Hot Gas Test Facility in Denver, Colorado, on a single tube cylindrical ESP using reentrained flyash from the Curtiss-Wright Small Turbine Test Rig.

The results include the evaluation of three corona electrodes in the 30.5 cm (1 ft.) diameter collector tube. The variables in the parametric tests were voltage, current, temperature, pressure and gas velocity. The effects of the various parameters on ESP performance are discussed.

INTRODUCTION

In January, 1985, a series of parametric tests on a bench scale, high temperature, high pressure electrostatic precipitator (HTHP-ESP) were concluded. In order to determine the optimum corona electrode for these tests, a series of electrode evaluation tests was performed. These tests were performed on three different corona electrodes operating at different power levels at a temperature of 900°C (1650°F), a pressure of 1 MPa (10 atm), and a gas velocity of 0.76 m/s (2.5 ft/s). Then final parametric evaluations were made on the ESP utilizing the optimum electrode design. These parametric evaluations were performed for temperatures of 843°C (1550°F) and 900°C (1650°F); pressures of 640 kPa (6.4 atm) and 1 MPa (10 atm); velocities of 0.76 m/s (2.5 ft/s) and 1.37 m/s (4.5 ft/s); electrode voltages of 85 kV and 115 kV; and corona currents of 5 mA and 10 mA.

DISCLAIMER

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

The tests were performed at the High Temperature High Pressure Electrostatic Precipitator (HTHP-ESP) test facility described by Rinard, et al. (1981) Figure 1 shows a schematic of the HTHP system. The HTHP-ESP is capable of operating at temperatures up to 980°C (1800°F) and pressures up to 1 MPa (10 atm) with a flow rate of 0.1 m³/sec (212 acfm) at these conditions. Collector tubes up to 35 cm (14 inches) in diameter can be tested in the unit. The collector tube used during the tests was 30.5 cm (12 inches) in diameter. The active section of the collector tube is 2.1 m (6.9 ft) long between inlet and outlet.

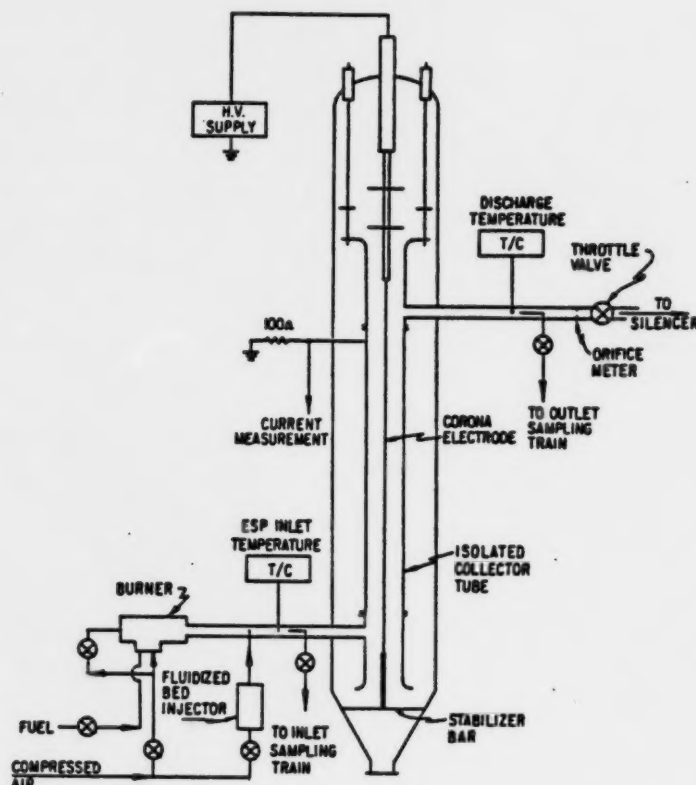


FIGURE 1. SCHEMATIC DIAGRAM OF HTHP ESP SYSTEM

TEST CONDITIONS

Fly ash from the second cyclone of the Curtiss-Wright PFBC system was reentrained into the gas stream. The elemental analysis of the ash is given in Table 1. Since the size distribution of the ash from the second cyclone did not provide sufficient particles in the range of 0.5 to 5 micrometers, the ash was ground in a fluid mill to reduce the particle size. The ground ash had a mass median diameter of 5.0 micrometer and a geometric standard deviation of 2.05. The ground dust was used in all of the efficiency tests.

TABLE 1. ELEMENTAL ANALYSIS OF CURTISS-WRIGHT ASH

SiO ₂	25.46
Al ₂ O ₃	12.72
TiO ₃	0.36
Fe ₂ O ₃	15.96
CaO	17.08
MgO	11.22
Na ₂ O	0.21
K ₂ O	0.77
P ₂ O ₅	0.10
SO ₃	18.12
TOTAL	102.00

Laboratory measurements of the resistivity of the ground fly ash from the second cyclone of the Curtiss-Wright PFBC were made in a parallel plate resistivity cell. A 0.44 cm thick layer of ash was placed between the plates, and a dc voltage was applied. The cross sectional area of the cell was 2.84 cm². The temperature of the cell was varied from 980°C (1800°F) to 316°C (600°F), and the applied voltage and the current through the ash were measured. The resistivity for this temperature range is given in Table 2.

TABLE 2. MEASURED FLY ASH RESISTIVITY

Temperature		Resistivity
°C	°F	ohm-cm
600	1112	6.8 x 10 ⁷
700	1292	9.0 x 10 ⁶
800	1472	1.2 x 10 ⁶
900	1652	2.6 x 10 ⁵
1000	1832	1.0 x 10 ⁵

Special precautions were made to document the operating temperature within the ESP. Since insertion of a thermocouple into the ESP tube would interfere with the electrical operation, temperature measurements during operation had to be made in the horizontal inlet section. To determine the temperature profile inside the ESP, the discharge electrode was replaced by a tube containing six shielded thermocouples along its length. The ESP was then operated over a range of temperatures and pressures, and the temperature inside the ESP was correlated with the measured inlet temperature and mass flow rate. This correlation was then used to determine the temperature of the ESP during operation. The temperature drop through the tube varied from 100-200°F with

the higher drops occurring at the higher temperatures and lower flow rates. The precipitator temperature that was reported during testing represents the temperature midway up the collector tube. The inlet temperature is also indicated.

In the corona electrode tests, the flow rate in the precipitator was maintained at $0.078 \text{ m}^3/\text{sec}$ (165 acfm) and the velocity through the collector tube was 1.07 m/sec (3.5 ft/sec). The collector tube had a specific collection area (SCA) of 25.6 sec/m ($130 \text{ ft}^2/\text{kacfm}$). The inlet gas temperature was measured, and the gas temperature in the ESP collector tube was calculated using the correlation function described above.

The three corona electrodes that were tested are shown in Figure 2. The first was a smooth wire 7.9 mm (5/16 inch) in diameter. The second corona electrode was a scalloped electrode designed by Research Cottrell (R-C) for use in a HTHP-ESP at the Curtiss-Wright PFBC facility (Feldman, 1977). Scalloped fins were attached to a 3.3 cm (1.3 inch) tube. This produced an electrode 8.3 cm (3.25 inches) in diameter. The third, or vane, electrode which was also designed by R-C; is similar to the scalloped electrode except that fins which are attached to the 3.3 cm (1.3 inch) tube are not scalloped.

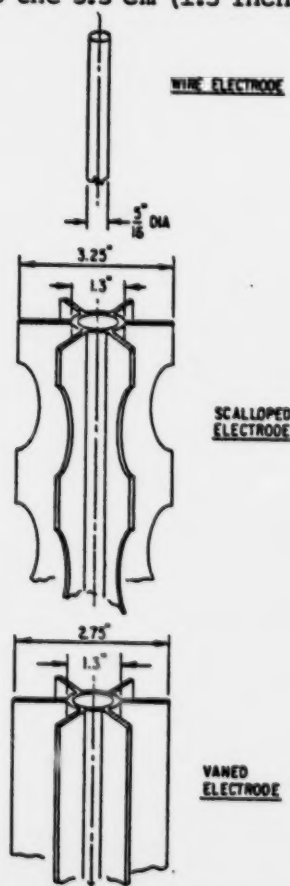


FIGURE 2. CORONA ELECTRODE

TEST RESULTS

Smooth Electrode Tests

In previous tests, the current increased dramatically at high temperatures when PFBC ash was present. A suspected cause for the high current was thermal emission of electrons from the ash surface. Before commencing parametric testing, tests were performed on a smooth wire electrode to gain more information about the cause of this excessive current. In these tests the V-I characteristics of a 1.59 cm (5/8 in.) diameter smooth electrode were measured under ash load conditions. Thermal emission would be indicated by: (1) current below corona onset; (2) current proportional to the relative areas and field strength at the surface of the negative electrode; (3) an abrupt increase of current at some temperature; and (4) negative ion charge carriers with either polarity voltage on the smooth electrode.

Figure 3 shows the negative and positive V-I curves for two dust conditions. The curves labeled "1" were recorded after the collector tube and electrode had been rapped and the system had been at temperature for about two hours without dust injection. The negative V-I curve appeared normal except for a small amount of current which was detectable at about 20 kV. The positive current was detectable at 10 kV and increased until 60 kV and remained reasonably constant until 120 kV. Below 130 kV, the positive current was larger than the negative current. The curves labelled "2" were recorded immediately after dust had been injected into the ESP for 15 minutes and then turned off. The addition of the dust caused the negative current to increase for voltages up to 120 kV and caused the positive current to increase abruptly for voltages above 10 kV and to remain larger than the negative current for the entire voltage range. The difference between the positive and negative

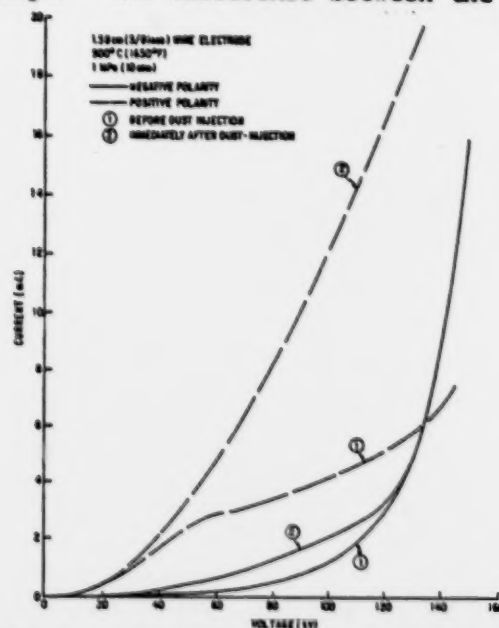


FIGURE 3. ELECTRICAL CHARACTERISTICS OF SMOOTH WIRE ELECTRODE

curves decreased, for both polarities, as the temperature decreased. However, the variation in the current as a function of temperature could not be determined since the dust conditions on the electrode could not be kept constant.

At a voltage of less than 100 kV, the current with positive polarity on the smooth electrode was approximately an order of magnitude larger than with negative polarity. Therefore, it did not appear that this current was due to either corona or thermal ionization. However, the current could be due to thermal emission from the cathode. An attempt to verify this was made by measuring the polarity of ion charge carriers. A charge-to-mass probe, developed by DRI (Durham, et al., 1972), was used to measure the polarity of charge carried by particles at the outlet sampling port. The polarity varied from positive to negative independently of voltage magnitude and polarity. It was concluded that triboelectric charging effects in the outlet duct masked the polarity of charge on the particles due to ion charge carriers. Without this data, the source of the currents at low voltage could not be determined.

Verification Tests

During the parametric evaluation, a series of verification tests were performed. The purpose of these tests was to insure that repeatable results were obtainable during the course of the testing. All verification tests were performed with the scalloped electrode. The operating temperature for the tests was 760°C (1400°F) since the highest mass collection efficiencies had been achieved at this temperature in previous tests. All but one of the verification tests were performed at 640 kPa (6.4 atm). During the mass measurements, the operating voltage was set just below sparkover and there was no rapping during the measurements.

The results of the verification tests are given in Table 3. The first verification test, V-1, was performed before the start of the Electrode Evaluation Tests to insure proper operation of the HTHP-ESP. The second verification test, V-2, was performed at the end of the Electrode Evaluation Tests to insure that the performance had not deteriorated since the initial verification test. The third verification test, V-4, was performed at this same time except at a pressure of 1 MPa (10 atm) to determine if the mass collection efficiency would remain high at the higher pressure. The final verification, V-4, was performed after the Final Parametric Evaluation was completed to determine if previously measured V-I characteristics and mass penetrations could be repeated.

As can be seen in Table 3, for all of the verification tests, penetrations of less than 1 percent were achieved. There was no deterioration in ESP performance between the first and last verification tests. The small differences in penetration measured during the verification tests is not significant since the accuracy of the measured penetration decreases for values of less than 1 percent due to the small amount of dust at the outlet. The electrical operating performance of the ESP was very stable and consistent throughout all verification tests. For the above reasons, it was concluded that the test procedures, instrumentation, and mechanical condition of the ESP were consistent throughout the test sequence.

TABLE 3. SUMMARY OF HTHP ESP VERIFICATION TEST RESULTS

TEST CONDITIONS: SCALLOPED ELECTRODE

T = 760°C, Vel = 0.76 m/s, SCA = 36 s/m
 (1400°F) (2.5 ft/s) (183 ft²/kacfm)

TEST NO.	V-1	V-2	V-3	V-4
NO. of RUNS	8	3	3	4
VOLTAGE (kV)	94	103	134	112
CURRENT (mA)	15	17	17	18
PENETRATION (%)	.73	.45	.94	.40
PRESSURE (psig)	82	82	135	82
FLOW RATE (ncfm)	215	214	337	216

Electrode Evaluation

The first step in the Parametric Tests Sequence of the Test Plan was the evaluation of the scalloped, wide-vaned, and smooth electrodes. These tests were all performed at 1 MPa (10 atm) and 900°C (1650°F). The gas velocity was 0.76 m/s (2.5 ft/s) and the SCA was 36 s/m (183 ft²/kacfm). The collector tube was rapped between mass concentration measurements.

The results for the scalloped electrode are presented in Table 4. During these tests, the corona current was set at four different levels: 40 mA, 20 mA, 10 mA, and 5 mA. For this electrode, it was not possible to maintain the same operating voltage for a given current. This can be seen from Table 4 by comparing the test runs 1.1a, b, and d'. For each of these tests, the voltage was 80 kV. However, the current varied from 39 mA to 5 mA for these three tests. It is further noted that the penetrations for these tests are all approximately 7.5 percent. Therefore, the penetration appeared to be much more a function of the voltage than the current. This is an important result since the power level varies by a factor of 8 for test 1.1a to 1.1d' with no change in penetration.

TABLE 4. SUMMARY OF HTHP ESP TEST RESULTS WITH SCALLOPED ELECTRODE

TEST CONDITIONS: SCALLOPED ELECTRODE AT VARIOUS POWER LEVELS

	1 MPa, (10 atm)	900°C, (1650°F)	0.76 m/s, (2.5 ft/s)	SCA = 36 s/m (183 ft ² /kacfm)			
TEST NO.	1.1a	1.1a'	1.1b	1.1b'	1.1c	1.1d	1.1d'
NO. OF RUNS	5	2	6	4	7	5	3
VOLTAGE (kV)	79	94	80	99	85	67	80
CURRENT (mA)	39	40	20	20	10	5	5
PENETRATION	7.5	3.4	7.7	5.2	5.8	11.5	7.2

Since the voltage varied very significantly during the tests, each test run was divided into primed and unprimed groups which corresponds to significantly different average operating voltage levels. As can be seen for all cases, the primed tests, which correspond to the higher average operating voltage for a given current, correspond to lower mass penetrations.

Table 5 gives a summary of the test results with the wide-vaned electrode. For the wide-vaned electrode, the operating voltage was more stable and did not vary in the manner described for the scalloped electrode above. Tests were performed at three power levels which corresponded to current levels of 17 mA, 10 mA, and 5 mA. In all cases, lower penetration was obtained for higher operating voltages which also corresponded to higher operating currents. The operating voltage at each of the current levels was higher than the scalloped electrode for corresponding current levels. However, the higher voltage did not result in a lower penetration for the wide-vaned electrode.

TABLE 5. SUMMARY OF HTHP ESP TEST RESULTS WITH WIDE VANED ELECTRODE

TEST CONDITIONS: WIDE VANED ELECTRODE AT VARIOUS POWER LEVELS

1 MPA, 900°C, 0.76 m/s, SCA = 36 s/m
(10 atm) (1650°F) (2.5 ft/s) (183 ft²/kacfm)

TEST NO.	1.2a	1.2b	1.2c
NO. of RUNS	3	3	3
VOLTAGE (kV)	129	114	82
CURRENT (mA)	17	10	5
PENETRATION (%)	4.6	8.1	15.2

A summary of test results for the smooth electrode is presented in Table 6. The operating voltage for the smooth electrode was stable as with the wide-vaned electrode. The significant feature of the smooth electrode is that it produces considerably less current for a given voltage than either of the other two electrodes. In addition, the smooth electrode was the only electrode that could be operated in a spark limiting condition. The largest variation in the electrical characteristics of the smooth electrode appeared to be in the sparkover voltage which varied from 110 kV to over 130 kV. Therefore, the smooth electrode could not be operated at higher voltage levels than the other electrodes since sparking would occur. When the ESP was operating in a sparking mode, penetration increased sharply as shown by Test 1.3a in Table 6.

TABLE 6. SUMMARY OF HTHP ESP TEST RESULTS WITH SMOOTH ELECTRODE

TEST CONDITIONS: 5/8" SMOOTH ELECTRODE AT VARIOUS POWER LEVELS

1 MPA, (10 atm)	900°C, (1650°F)	0.76 m/s, (2.5 ft/s)	SCA = 36 s/m (183 ft ² /kacfm)
TEST NO.	1.3a	1.3b	1.3c
NO. of RUNS	1	4	3
VOLTAGE (kV)	126*	100	97
CURRENT (mA)	10	5	2.5
PENETRATION (%)	21.7	10.1	13.0

In Figure 4, penetration is plotted as a function of voltage for the three different electrodes operating at various voltage levels as shown in Tables 4, 5 and 6. For each of the electrodes, penetration decreased as the voltage was increased. Although the wide-vented electrode could be operated at the highest voltage levels, the higher voltage did not result in lower penetration as would be expected. For a given voltage, the penetration with the scalloped electrode was less than either of the other two electrodes which both performed at about the same collection efficiency for a given voltage. The reason for this may be that the current distribution for the scalloped electrode is more uniform than the other two electrodes. Penetration determines the ESP size and the power density determines the operating cost of the ESP which has been sized to achieve a specific penetration. For each test, the values of migration velocity, w_k , and power density, P_D , were calculated and plotted in Figure 5. For each test, the SCA required to achieve one percent penetration, as determined by the modified Deutsch equation, is given on the vertical scale on the right-hand side. The required power consumption per unit volume of gas treated is shown by the family of straight lines that pass through the origin. The steeper the line, the lower the power consumption per unit gas flow, and therefore, the lower the operating cost. The capital cost increases with SCA. Therefore, the most cost effective operating region is in the upper left hand corner of the diagram.

Figure 5 shows that the lowest SCA and the lowest power consumption were achieved with the scalloped electrode. It appears that the most cost-effective electrical operating range was at a voltage of 80 kV to 85 kV where current densities were 2.5 to 5 mA/m². In this range, power consumption of 8.5 to 17 kW-s/m³ (4 to 8 W/scfm) of gas treated and SCAs of 95 to 108 s/m (480 to 550 ft²/kacfm) would be required to achieve a penetration of 1% or a collection efficiency of 99%. If these projected values of size and power consumption are valid, then the use of an ESP as a hot gas cleanup system at 900°C (1650°F) and 1 MPa (10 atm) would be technically feasible. Based on these results, no additional electrode designs were evaluated, and the scalloped electrode was selected for final parametric evaluation.

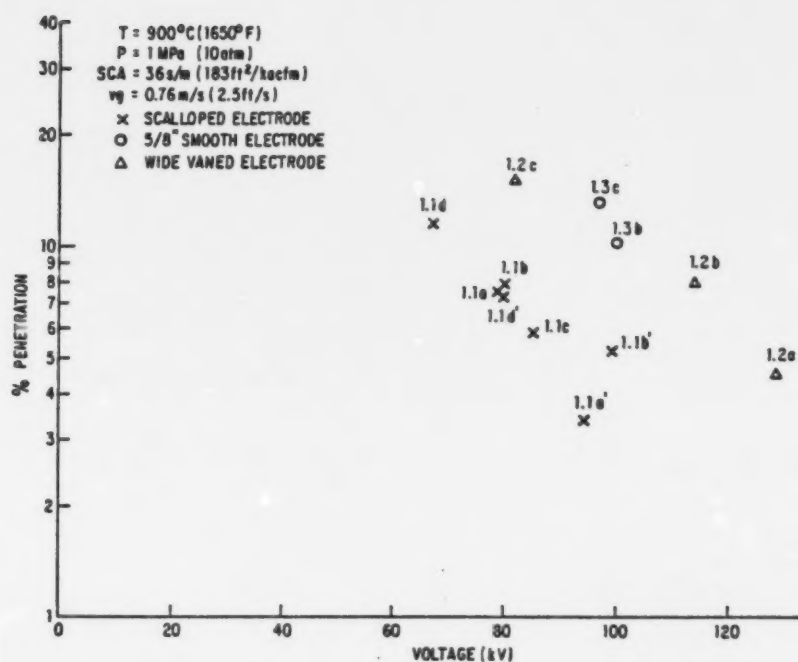


FIGURE 4. PENETRATION VS VOLTAGE FOR SCALLOPED, WIDE VANED AND SMOOTH ELECTRODE

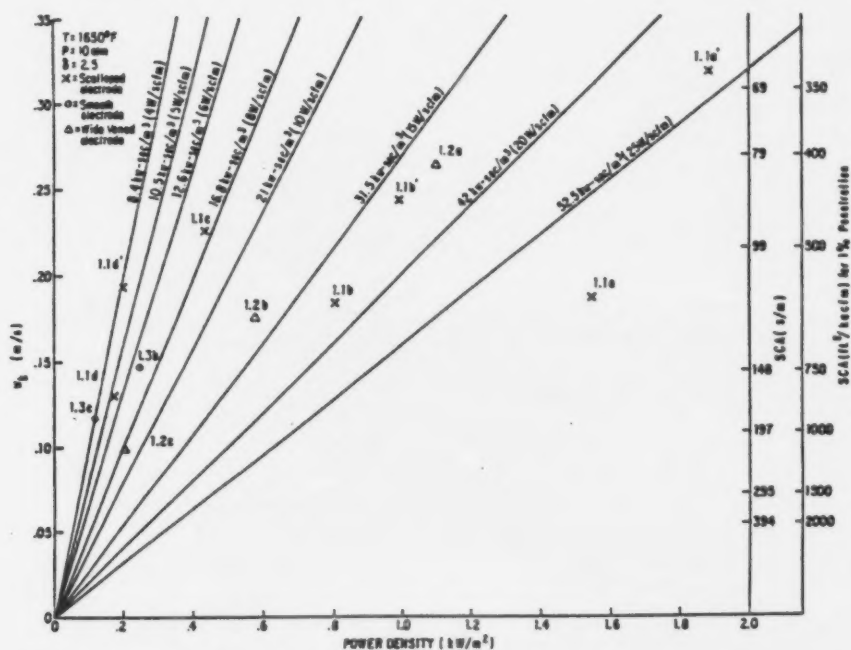


FIGURE 5. POWER CONSUMPTION AND SCA TO ACHIEVE 1% PENETRATION

Final Parametric Evaluation

The tests in the Final Parametric Evaluation were to determine the sensitivity of ESP performance to changes in pressure, temperature, voltage, current and gas velocity. The scalloped electrode which was selected as the best performing electrode in the Electrode Evaluation tests was used for these tests. At 900°C (1650°F) and 1MPa (10 atm), the nominal operating point which appeared to be most cost effective was a voltage of about 85kV and a current of about 10mA (5mA/m²).

Four of the parametric tests were performed with a constant gas velocity of 0.76m/s (2.5 ft/s) in the 30.5cm (1 ft) diameter collector tube. V-I curves showing the typical electrical characteristics during the tests are plotted in Figure 6. Also, the average voltage and average current for all the runs during each test are shown in this figure. Test 5.1a was a repeat of Test 1.1c which was at the nominal operating point determined during the Electrode Evaluation. However, for a current of about 10mA, the voltage in Test 5.1a was 116kV compared to 85kV in Test 1.1c. This change in V-I characteristics followed the same trend that was observed earlier where the voltage was not constant for a given value of current. The higher voltage resulted in a lower penetration as shown in Table 7 where for Test 5.1a the penetration was 3.3% compared to 5.8% shown for Test 1.1c in Table 4.

TABLE 7. SUMMARY OF HTHP ESP TEST RESULTS FOR PARAMETRIC EVALUATION

TEST CONDITIONS: Scalloped Electrode -

Variables are pressure, temperature and velocity.

TEST NO.	5.1a	5.1b	5.2	5.3	5.4	5.5
NO. OF RUNS	6	5	7	7	5	6
VOLTAGE (kV)	116	90	89	115	84	91
CURRENT (mA)	9	4.8	10	5.9	10	5
PENETRATION (%)	3.3	9.1	5.6	3.2	15.4	8.4
ESP TEMP (°F)	1655	1660	1650	1555	1655	1545
PRESSURE (psig)	135	135	82	135	135	135
FLOW RATE (ncfm)	292	295	188	308	528	558
VELOCITY (ft/sec)	2.5	2.5	2.5	2.5	4.5	4.5

In Test 5.2, the temperature was maintained at 900°C (1650°F) and the pressure was reduced from 1MPa (10 atm) to 640kPa (6.4 atm) which resulted in the change in V-I characteristics as shown in Figure 6. For a constant current level of about 10mA, the reduction in pressure was accompanied by a reduction in voltage from 116kV in Test 5.1a to 89kV in Test 5.2 (due to the lower gas density). A comparison of the performance measurements in Table 7 shows that the penetration was 5.6% during Test 5.2 and 3.3% during Test 5.1a. This indicates that for a constant temperature and current, a reduction in pressure results in a decrease in voltage and an increase in penetration.

To determine if the increase in penetration was due to the decrease in voltage or was also a function of the pressure change, Test 5.1b was performed where, in respect to Test 5.2, the temperature and voltage were kept constant at 900°C (1650°F) and about 90kV and the pressure was increased from 640kPa (6.4 atm) to 1MPa (10 atm) resulting in a current decrease from 10mA to 4.8mA. The penetration increased from 5.6% in Test 5.2 to 9.1% in Test 5.1b. This result indicated that penetration cannot be predicted by considering only the voltage level when additional parameters are changed. A comparison of the result from the Electrode Evaluation Test 1.1c, at 1MPa (10 atm) in which the voltage was 85kV and the current was 10mA, with Test 5.2 at 640kPa (6.4 atm), in which the voltage and current were 89kV and 10mA, shows the penetrations were essentially the same for the two different pressures when both voltage and current were unchanged.

It appears that an increase in pressure does not effect ESP performance if neither voltage nor current are changed. However, when pressure is increased an ESP can be operated at a higher voltage level which results in a lower penetration as shown by Test 5.1a and Test 5.2. Also, these two tests at 900°C (1650°F) show that the power per standard volume of gas required to achieve 1% penetration at 1MPa (10 atm) is approximately half that at 640kPa (6.4 atm).

Test 5.3 was performed at 845°C (1550°F) to determine the effects of temperature on ESP penetration when pressure and either voltage or current are held constant as shown in Figure 6. Table 7 shows that the penetrations during Test 5.1a and Test 5.3 were essentially the same when pressure and voltage were held constant at 1MPa (10 atm) and 115kV and the temperature decreased from 900°C (1650°F) to 845°C (1550°F) and the current decreased from 9mA to 4.8mA. These two tests did not show the large decrease in penetration at 845°C (1550°F) that was observed during the verification tests at 760°C (1400°F). The probable reason for this fact is that the voltage was lower and the temperature higher in Test 5.3 as compared to the verification tests. Test 5.1a and Test 5.3 showed that the power consumption per standard volume of gas to achieve 1% penetration would be about 13.5kW^{-s}/Nm³ (6.4 W/scfm) at 900°C (1650°F) and about 8.3kW^{-s}/Nm³ (3.9 W/scfm) and the required SCA would be about 65s/m (330 ft²/kacfm) for both conditions.

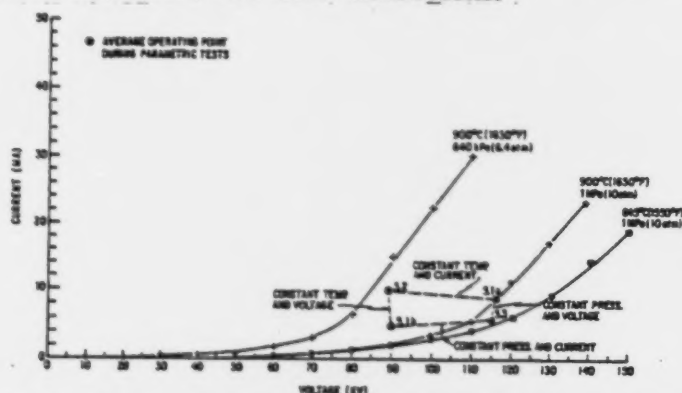


FIGURE 6. VI CURVES AND OPERATING POINTS DURING PARAMETRIC TESTS AT 0.76 m/sec (4.5 ft/sec).

A comparison of Test 5.3 with Test 5.1b shows that at a constant pressure and current level that as the temperature is increased the voltage decreases and the penetration increases. This characteristic has been observed consistently throughout the ESP tests.

Test 5.4 and Test 5.5 were performed with a gas velocity of 1.37m/s (4.5 ft/s) and an SCA of 20s/m (101 ft²/kacfm) for comparisons with Test 5.1a and Test 5.3 where the gas velocity was 0.76m/s and the SCA was 36s/m (183 ft²/kacfm). It was planned that temperature, pressure, voltage and current would be kept constant and only gas velocity would be changed. However, when the gas velocity was changed, the electrical characteristics changed as shown in Figure 7 and Figure 8. At 1MPa (10 atm) and 900°C (1650°F), Figure 7 shows that the operating voltage level in Test 5.4 was 84kV with a gas velocity of 1.37m/s (4.5 ft/s) compared to 116kV at a gas velocity of 0.76m/s (2.5 ft/s) in Test 5.1a. Figure 8 shows that at 845°C (1550°F), the increase in gas velocity caused a comparable change in the voltage level. The main reason for the change in electrical characteristics with an increase in gas velocity is probably due to the changes in temperature gradients along the ESP collector tube. Although an attempt was made to keep the temperature constant at the mid-length of the collector tube by adjustment of the inlet temperature for each velocity, the effective temperature of the system appeared to increase as gas velocity increased.

Table 7 shows that the penetration increased from 3.3% to 15.4% when the velocity was increased from 0.76m/s (2.5 ft/s) in Test 5.1a to 1.37m/s (4.5 ft/s) in Test 5.4. At 845°C (1550°F), the penetration increased from 3.2% in Test 5.3 to 8.4% in Test 5.5 when the velocity was increased. In both cases, the increase in penetration was due to the decrease in voltage and SCA and perhaps the increase in velocity.

To compare the performance of the ESP at the two gas velocities, the modified migration velocity (w_k) which is approximately independent of SCA was calculated using Equation (1)

$$w_k = (\ln p)^2 / \text{SCA} \quad \text{m/s} \quad (1)$$

Where w_k = modified migration velocity
 p = penetration
 SCA = Specific Collection Area

These results are tabulated in Table 8. Results for Test 1.1c are also included in this table. At 900°C (1650°F), w_k decreased significantly at the higher velocity indicating that the increase in penetration was due to other factors in addition to the decrease in SCA. For Test 1.1c, in which the voltage and current levels were the same as in Test 5.4 but the velocity was 0.76m/s (2.5 ft/s), w_k was 0.23m/s which was larger than w_k at the higher velocity in Test 5.4. This indicated that in addition to the lower voltage, the increase in gas velocity caused a decrease in ESP performance at 900°C (1650°F). However, this was not the case at 845°C (1550°F) since w_k was

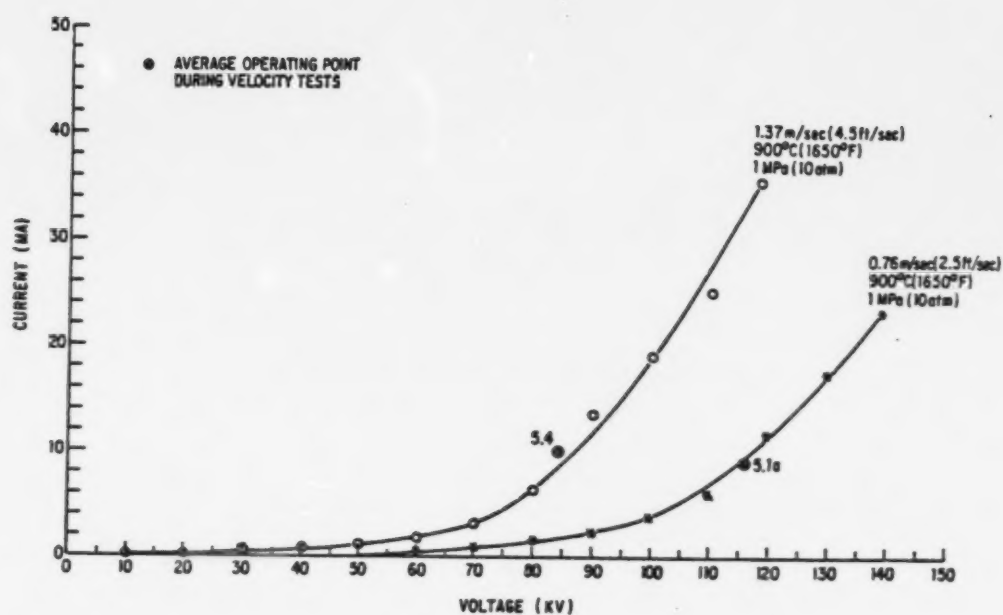


FIGURE 7. VI CURVES AND OPERATING POINTS DURING VELOCITY TESTS AT 900°C (1650°F)

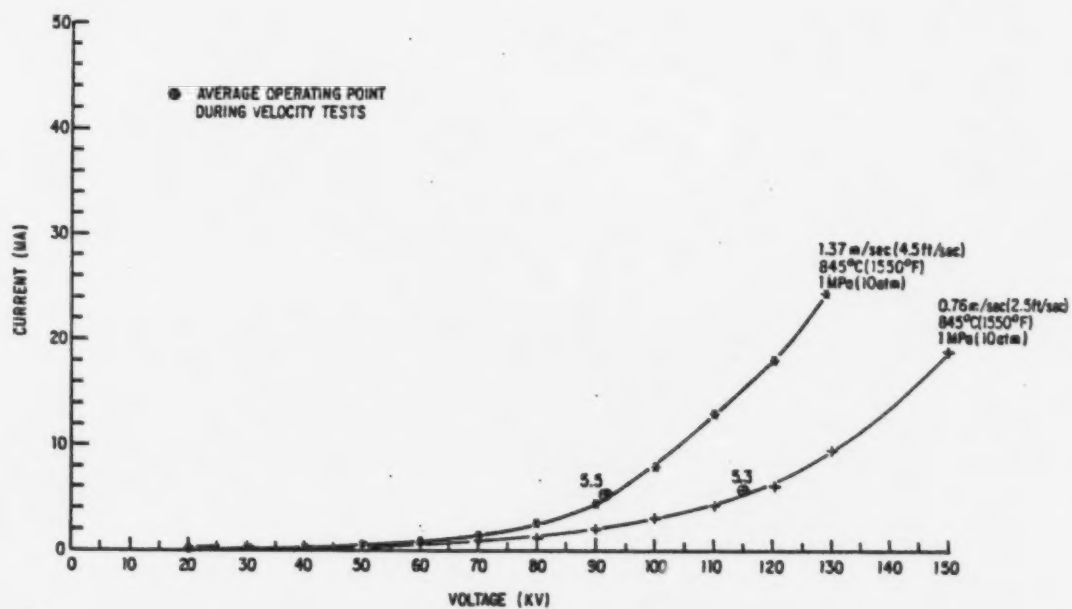


FIGURE 8. VI CURVES AND OPERATING POINTS DURING VELOCITY TESTS AT 845°C (1550°F)

essentially the same for both velocities as shown in Table 8. This indicates that at 845°C (1550°F), the performance of an ESP with a gas velocity of 1.37 m/s (4.5 ft/s) would be as good as the performance at 0.76m/s (2.5 ft/s) provided the SCA was the same.

TABLE 8. MODIFIED MIGRATION VELOCITIES FOR GAS VELOCITIES OF
0.76m/s (2.5ft/s) 1.37m/s 4.5ft/s)
AT A PRESSURE OF 1 MPa (10 atm)

Test	Temperature	Velocity	w_k
5.1a	1650°F	2.5 ft/s	0.32m/s
5.4	1650°F	4.5 ft/s	0.18m/s
5.3	1550°F	2.5 ft/s	0.33m/s
5.5	1550°F	4.5 ft/s	0.31m/s
1.1c	1650°F	2.5 ft/s	0.23m/s

CONCLUDING REMARKS

The verification tests indicated that the ESP performance, instrumentation and sampling procedures were consistent and reliable throughout the entire test sequence. The repeatability of test results during verification tests further demonstrates stable operation. Previous test results (Rinard, et al., 1984) have shown the bench scale ESP to give results consistent with a pilot scale ESP on an operational PFBC. For these reasons, high confidence may be placed in the parametric test results.

The test results indicated that the scalloped electrode provided the highest collection efficiency and lowest power consumption. Furthermore, if the scale-up assumptions are valid, the scalloped electrode would make hot gas cleanup by use of an ESP at 900°C (1650°F) and 1 MPa (10 atm) technically and economically feasible. However, the scalloped electrode did exhibit a wide variation in voltage-current characteristics, and there is no way, at this time, to predict that the scalloped electrode will maintain a prescribed operating point. The increased current which sometimes occurred at high temperature could be due to thermionic emission of electrons from the ash surface. Further work is required to investigate this phenomenon.

In all cases, for a given electrode, the highest performance was obtained with the highest operating voltage. It has generally been assumed that high current density is required for high temperature (low ash resistivity) ESP operation (Feldman, Kumar, 1982). However, the results obtained here do not support this assumption. The same collection efficiency was obtained with the scalloped electrode when operating at the same voltage even though the current was nearly an order of magnitude different. However, when electrodes which produced less current were tested at higher voltage levels, performance did not improve. The reason for the lack of improvement may have been that the electrodes produced a non-uniform current distribution. These facts indicate that further work is required to determine the cause for the wide variation in

electrical characteristics in order that an electrode can be designed that is stable and operates at high voltage (for good collection) and a low uniform current (for low power consumption). To date, this issue has not been adequately resolved. The resolution of this issue is crucial to the application of ESPs to PFBC and other high temperature high pressure applications.

ACKNOWLEDGMENTS

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DESIGN OF BAGHOUSE AND ELECTROSTATIC PRECIPITATOR
FOR THE COAL-FIRED FLOW FACILITY*

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ABSTRACT

Design studies for a baghouse and an electrostatic precipitator that have been installed in a pilot-scale coal-fired MHD flow train are discussed and a description of the equipment is given. Results of initial tests are also discussed.

INTRODUCTION

The University of Tennessee Space Institute (UTSI), under contract with the Department of Energy, is obtaining data required for design of components for coal-fired power plants equipped with a magnetohydrodynamic (MHD) topping cycle. A pilot MHD flow train, referred to as the Low Mass Flow (LMF) train is being tested at the DOE Coal Fired Flow Facility (CFFF) located on the UTSI campus. The LMF train simulates operation of components in an MHD power plant. A baghouse and an electrostatic precipitator have been installed in the LMF train and are being operated simultaneously to determine the suitability of these devices for controlling particulate emissions and to provide design data for future large scale installations.

Due to very high combustion temperatures and addition of potassium carbonate "seed" to the coal feed, both required to increase the electrical conductivity of the flue gas, MHD particulates have a mass mean diameter of about 1 μm and consist of about 85 percent potassium sulfate resulting from combination of the potassium seed with sulfur dioxide. Due to the unique properties of the dust, design studies were undertaken for both the baghouse and the ESP. These studies will be discussed in the paper. Results of

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initial tests conducted during 1985 will also be discussed.

THE MHD TOPPING CYCLE

The MHD topping cycle involves firing coal, oil or gas seeded with a metal salt at temperatures of about 5000°F to generate an electrically conductive plasma. The plasma, contained in a channel made up of electrodes, is passed through a magnetic field. Motion of the plasma through the magnetic field converts some of the kinetic energy into electricity by the same process that causes current flow in the windings of a generator. Once the plasma has been cooled to about 3000°F the electrical conductivity is too low for efficient MHD power generation, so a more-or-less standard steam generating system is used to recover the remaining energy from the gas. Theoretical studies have shown that overall conversion efficiencies as high as 55% can be reached using the combined MHD-steam cycle, compared to efficiencies of less than 40% for the steam cycle alone.

THE LOW MASS FLOW TRAIN

The Low Mass Flow train, shown in Figure 1, is a 23 MW_e nominal system that fires a combination of coal and fuel oil. An oil-fired vitiation heater preheats oxidizer to about 1700°F. A mixture of coal and oil is burned in a pressured combustor at about 5000°F with stoichiometry of about 0.85. The hot gas is accelerated through a supersonic nozzle and passes through an MHD channel. The channel is made up of segmented copper electrodes and is surrounded by a magnet. Power is removed by leads connected to the electrodes.

A radiant furnace is used to provide a relatively slow gas cooling rate to allow NO_x formed during combustion to decompose. About 40% of the total solids is tapped at the bottom of the furnace. Following the radiant furnace is a secondary combustor that introduces air to burn the remaining CO and H₂ in the flue gas. The final stoichiometry is about 1.05.

A Superheater Test Module (SHTM), consisting of a series of water and steam cooled tube banks in a horizontal duct, is used to simulate gas-side conditions in the steam generating system in order to obtain heat transfer, fouling, and corrosion data. Steam cooled sections simulate operation of secondary superheater, intermediate temperature air heater, and primary superheater surface. Water cooled sections are used to reduce gas temperatures to the desired level for the three test sections. The flue gas is cooled from about 2300°F to about 800°F in the SHTM. A recuperative air heater is used to preheat the air supplied to the secondary combustor from ambient to about 600°F.

Particulate collection devices including a venturi scrubber-cyclone system, a baghouse, and an electrostatic precipitator are located in parallel streams at the exit of the SHTM. As shown in Figure 2, the ductwork is arranged so set gas flow rates can be maintained in the baghouse and ESP, while the excess gas passes through the venturi scrubber-cyclone system. The

baghouse and precipitator are both sized for a maximum flow rate of 8500 cfm, which is 50% of the total LMF train flow at nominal conditions. Gas temperature at the entrance of the particulate collection devices is about 400°F. Descriptions of the baghouse and ESP are given in later sections.

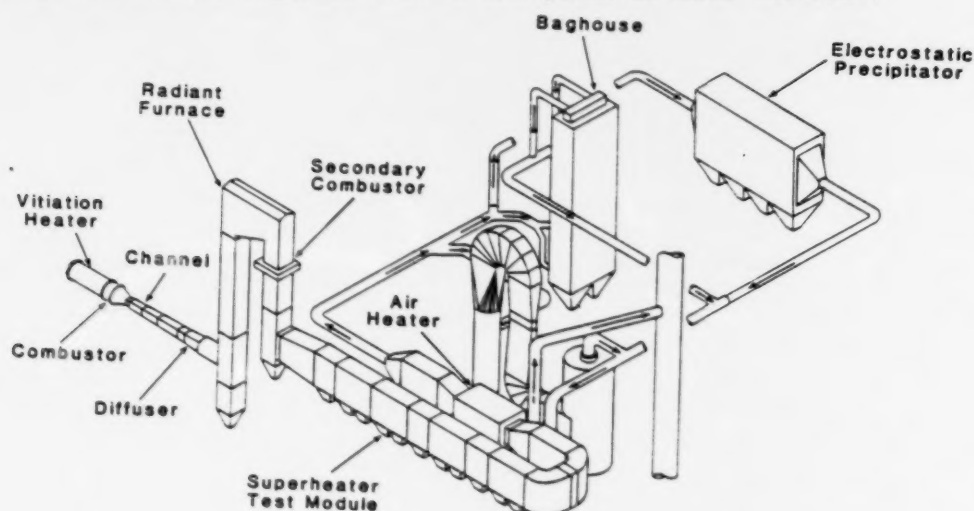


Figure 1. Low Mass Flow Train

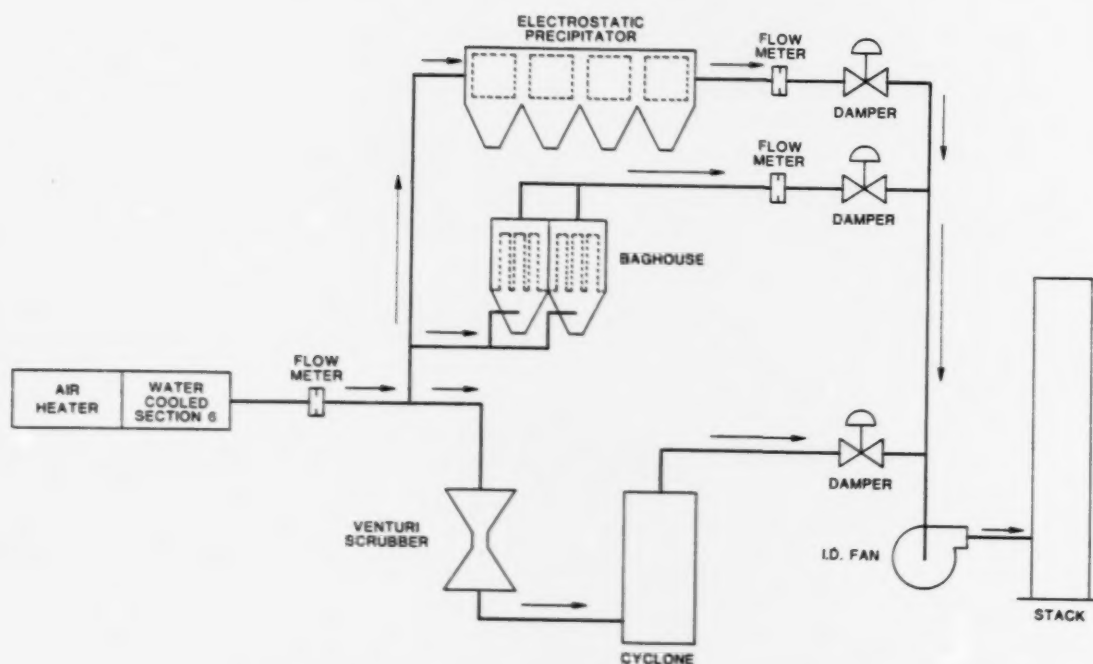


Figure 2. LMF Downstream Flow Schematic

MHD FLY ASH CHARACTERISTICS

Due to the very high initial combustion temperature, MHD fly ash is formed mostly by condensation of vaporized potassium and ash constituents from the coal. If the sulfur content of the coal is high enough the potassium seed, which makes up about 10% by weight of the coal feed as K_2CO_3 , shows up as K_2SO_4 in the fly ash. If not, the excess forms K_2CO_3 .² In testing at the CFFF to date a high sulfur Illinois No. 6 coal has been burned, so the K_2CO_3 content of the ash has been negligible. Since K_2SO_4 has a lower boiling point than most of the compounds formed by the other ash constituents, only a small fraction of the potassium is tied up as slag and the fly ash is typically 85 to 90% K_2SO_4 . A typical coal analysis is given in Table 1 and typical coal ash, slag and fly ash analyses are given in Table 2.

Table 3 shows a range of typical ash dispositions in the system. Slag tapped in the radiant furnace ranges from about 30% to about 50% of the ash input, while fly ash at the SHTM exit ranges from 35 to 50%. These variations are apparently due to changes in process conditions from test to test. It is interesting to note that the percentage of ash deposited in the SHTM and the fly ash loading at the SHTM exit are very similar for all tests. The estimated particulate collection efficiency of 99.0 to 99.2% required to meet the NSPS limit of 0.03 lb/10⁶ Btu is based on tests of less than 20 hours duration. It is expected that ash deposits in the SHTM, now about 17% of the total, will be reduced once equilibrium is established during longer tests. Since the fly ash loading may be higher in future tests, a target efficiency of 99.6% has been established for the particulate collection devices.

TABLE 1. TYPICAL SEEDED COAL ANALYSIS (% BY WEIGHT)

H ₂ O	3.2
ASH	8.7
SEED	10.2
S (Organic)	1.7
C	58.7
H	3.9
N	1.3
O	12.4

TABLE 2. TYPICAL SEEDED COAL ANALYSIS (% BY WEIGHT)

	<u>Coal Ash</u>	<u>Tapped Slag</u>	<u>Fly Ash</u>
SiO ₂	44.9	40.7	4.5
Al ₂ O ₃	16.6	23.5	0.8
Fe ₂ O ₃	13.6	14.3	3.0
TiO ₂	1.0	0.8	0.3
CaO	3.6	3.6	0.4
MgO	0.5	0.6	0.1
Na ₂ O	0.3	0.1	0.5
K ₂ O	9.6	16.1	48.4
SO ₃	9.8	0.4	42.2

TABLE 3. DISPOSITION OF ASH IN LMF TRAIN

Test	Tapped Slag %	SHIM Deposits (By Difference) %	Fly Ash At SHIM Exit, %	Fly Ash Flow (Lb/Min)	Fly Ash Loading (GR/DSCF)	Fly Ash Loading (GR/ACF)	Collection Efficiency Required to Meet NSPS, %
LMF3-B1	32.4	17.5	50.1	5.05	3.40	1.35	99.2
LMF4-B	48.3	16.7	35.0	3.87	3.43	1.66	99.0
LMF4-C	41.3	16.8	41.9	4.31	3.38	1.77	99.1

Since the fly ash condenses from vapor, the particles produced are very small. The mass mean particle size, determined by Brink impactor tests, is typically on the order of 1 μm . A representative size distribution is shown in Figure 3.

The collected particulate has a low bulk density, typically about 20 lb/ft³. The particles have a strong tendency to adhere to each other and to surfaces they contact, so formation of thick dust cakes in the dust collectors and bridging in hoppers are likely. K_2CO_3 and to a much lesser extent K_2SO_4 are hygroscopic, so the dust properties are strongly affected by the moisture content of the flue gas.

An additional property of the dust which is important in operation of electrostatic precipitators is electrical resistivity. Resistivities of samples collected during LMF testing are shown in Figure 4. The resistivity of a sample containing a significant fraction of K_2CO_3 was much lower than that of samples containing mostly K_2SO_4 , but in all cases the resistivity was lower than that of a typical "low resistivity" fly ash.

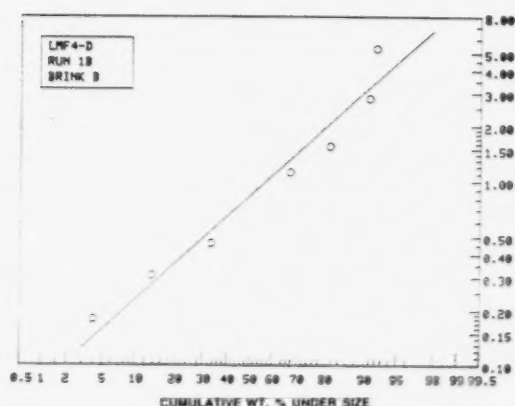


Figure 3. Typical Particle Size Distribution

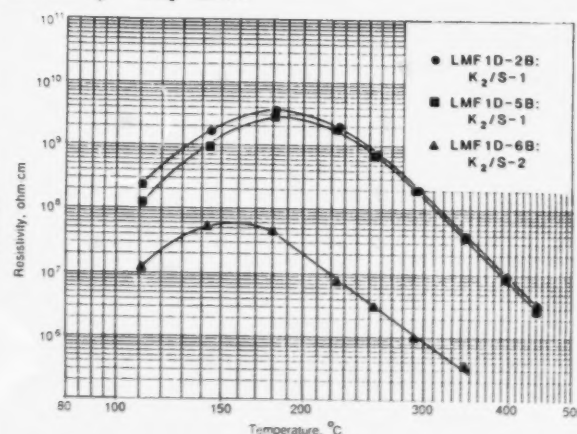


Figure 4. Fly Ash Resistivity As A Function of Temperature

APPLICATION OF BAGHOUSE AND ELECTROSTATIC PRECIPITATOR FOR MHD PARTICLE REMOVAL

Baghouses and electrostatic precipitators are the two commonly used devices for removing particles from utility boiler flue gas. Due to increasingly strict air pollution regulations, a number of baghouses have been installed in recent years. However, the latest economic survey showed that ESP's are still economical for some applications with the choice depending on the fly ash properties (1). Since there is no previous operating experience for baghouses collecting MHD ash and only very limited pilot plant data is available for ESP's, it is not possible to make a technical or economic comparison between the two devices at present. Baghouse and ESP testing at the CFFF should provide the data necessary for these comparisons.

MHD ash bears little resemblance to fly ash produced by coal-fired boilers. The low odor kraft recovery boiler operation is the industrial process that produces dust that is most similar to MHD dust (2), (3), (4). In this process spent liquor from the kraft pulping process used to make paper is burned to recover inorganic chemicals for reuse and to generate steam for use in the process. The dust that is formed by burning the liquor consists mostly of Na_2SO_4 with smaller amounts of Na_2CO_3 and NaCl . The size of the dust ranges from 50 to 85 weight percent less than $2\text{ }\mu\text{m}$ diameter and the loading ranges from 4 to 11 gr/dscf. Historically, electrostatic precipitators have been used to collect dust from the kraft recovery process and this operating experience was heavily relied on in the CFFF precipitator design. Baghouses are not used on kraft recovery operations because the dust is often sticky due to adsorption of SO_3 and H_2SO_4 on dust particles. It is not known if a similar process will occur with MHD dust; however, as the SO_3 content of the MHD flue gas is very low it seems unlikely. One would suspect that the small particle size is an additional reason why baghouses are not used on kraft recovery boilers.

Due to the unique properties of MHD fly ash a design study was undertaken for both the baghouse and the ESP. Details of these studies are given in the following sections.

BAGHOUSE DESIGN STUDY

Two dust properties were identified as likely sources of baghouse operating problems. The first of these is the hygroscopic nature of the dust, particularly when significant amounts of K_2CO_3 are present. It was decided that there is no practical way to keep a baghouse warm enough to prevent adsorption of water during down periods, but that the problem could be minimized by passing through the water dew point as quickly as possible at startup and by minimizing the amount of dust remaining on the bags and in the hoppers after shutdown.

The second dust property that was identified as a potential source of operating problems was the small particle size. Small particles can become embedded in the fabric so they can not be removed by the cleaning system.

This problem, known as "blinding", leads to unacceptably high pressure drops across the bags. Another problem related to particle size is settling time. Fine dust is easily reentrained in the gas stream after cleaning.

Based on the above considerations, the baghouse design is determined by choosing a cleaning method and a bag material that minimize the possibility of blinding and dust reentrainment. The four commonly used cleaning methods are on-line pulse jet, off-line pulse jet, reverse gas, and shake-deflate. Any type of on-line cleaning was ruled out, since the dust would not settle into hoppers. Off-line pulse jet cleaning was not selected because this system is not typical of utility baghouses. That pulse jets are not common seems to be due to the comparatively large number of bags required and the extensive piping system required for the compressed air. The latest EPRI compilation of utility baghouse data (5) lists a total of 84 operating baghouses. Of this number 71 were reverse gas, 10 were shake-deflate and 3 were pulse jet. The shake-deflate cleaning method is promising for utility applications because it provides more efficient cleaning than the reverse gas method, thus allowing operation at higher air/cloth ratios. However, the reverse gas system is still the preferred cleaning method and much of the advantage in cleaning efficiency provided by the shaker method can be obtained by using sonic horns along with the reverse gas. It was therefore decided to use the reverse gas cleaning system, with provision for sonic horn assistance, in the CFFF installation.

The three commercially available bag materials that can operate above 300°F are fiberglass, Teflon and Nomex. Nomex is not used on boiler baghouses because it is deteriorated by SO_2 . Excluding pulse jet units woven fiberglass bags, usually with a 10% Teflon coating, are used in all utility baghouses. In order to determine a suitable bag material for collecting MHD dust, tests of four different fiberglass filter media were conducted in a small bench scale dust collector at the University of Tennessee in Knoxville (6). The four fabrics tested were a 14.2 oz/yd² 3x1 twill, 44x24 (WxF) thread count woven fiberglass fabric with a Teflon B finish (Fabric Filter Co. #502-1TC); a 16.8 oz/yd² 50x30 (WxF) thread count woven fiberglass fabric with a special filament yarn and a Teflon B finish (Fabric Filter Co. #SP-1511); an experimental fiberglass media for use in reverse gas applications with a simulated felted surface known as Huyglas, 16 oz/yd² with a Teflon finish (Huyck Felt Co.); and a 9.8 oz/yd² 3x1 twill weave fiberglass media with a Gore-Tex membrane on the filtration surface (W.L. Gore and Associates). An iron oxide dust with a mass mean diameter of 1.9 μm was used to simulate MHD dust. Each fabric was tested at a filtration velocity of 3 ft/min with a dust loading of 3 gr/ft³. The fabrics were conditioned by repeated 30 minute cycles and were cleaned between each cycle by a small pulse of air simulating reverse gas cleaning. Results of the tests are shown in Figures 5, 6 and 7. Figure 5 is a plot of residual pressure drop vs. filtration cycle number. Figure 6 is a plot of pressure drop at 30 minutes into the cycle vs. cycle number. Figure 7 is a plot of pressure drop vs. filtration time during the 21st cycle. The Huyglas and Gore-Tex fabrics seem to have conditioned normally, while the residual and maximum pressure drops of the woven fabrics continued to rise. In Figure 7 it can be seen that the rate of increase in pressure drop approached a constant value for all four fabrics, but that the initial pressure drops were significantly different.

Observations showed that all fabrics except the Gore-Tex retained some dust cake after cleaning, while the Gore-Tex fabric had a clean filtration surface. Collection efficiency for the woven fabrics was 99.2 to 99.8% while collection efficiency for the Huyglas and Gore-Tex fabrics were over 99.9%. As a result of the tests it was determined that woven fiberglass bags would probably not perform well collecting MHD dust, but that the Gore-Tex fabric has a good chance of performing acceptably.

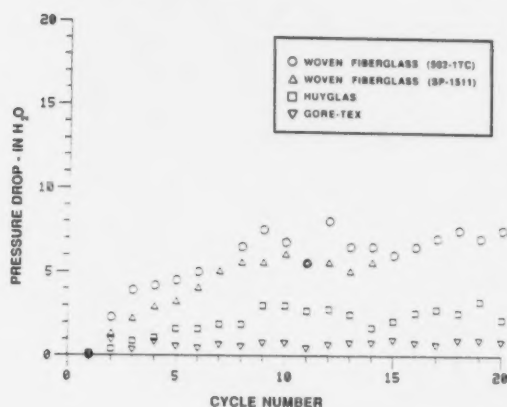


Figure 5. Residual ΔP Vs. Cycle Number-Bench Scale Test

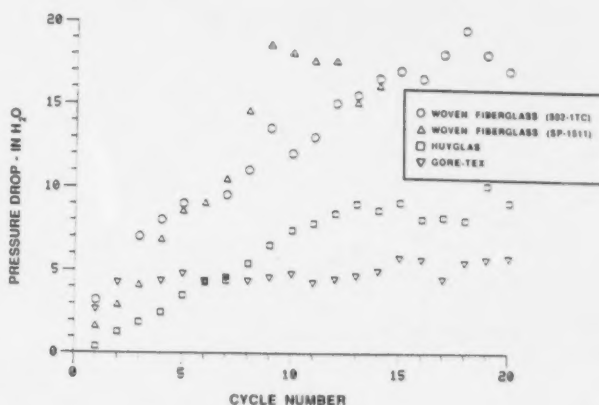


Figure 6. Maximum ΔP Vs. Cycle Number-Bench Scale Test

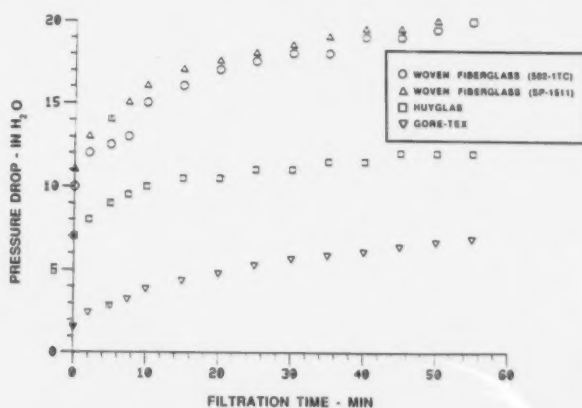


Figure 7. ΔP Vs. Filtration Time-Bench Scale Test

It was decided to design the baghouse for a maximum filtration velocity of 3.0 ft/min, with a range of 1.5 to 3.0 ft/min to cover the commonly used range. 11.5 inch diameter by 30 foot long bags were used because they are typical of utility baghouse installations. For a maximum gas flow rate of 8500 cfm and a maximum filtration velocity of 3 ft/min, the number of bags was set at 30. The baghouse was divided into two compartments to provide a source of clean hot gas for cleaning. During cleaning part of the normal gas flow is diverted to the venturi scrubber.

The baghouse, shown in Figure 8, is constructed from 3/16 inch carbon steel plate and has 4 inches of external insulation. It is 6.5 ft x 11.75 ft in cross section and 54 ft tall. Each compartment contains 3 rows of 5 bags suspended open end down. Reverse gas and outlet poppet dampers are located at the top of each compartment. The reverse gas fan is rated at 2150 cfm. Sonic horns are mounted below the top walkway about 8 feet below the roof. These horns produce sound at a basic frequency of 125 Hz at an intensity of 143 dB.

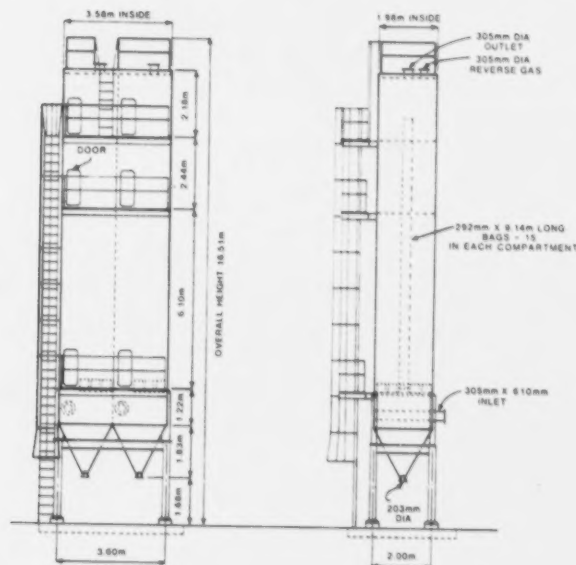


Figure 8. LMF Baghouse

INITIAL BAGHOUSE TEST RESULTS

The baghouse was operated during four tests in the summer and fall of 1985 for a total of about 22 hours. Table 4 gives a summary of the residual pressure drops, pressure drops 30 minutes into the filtration cycle, and maximum pressure drops for each cycle during these tests. The nominal flow rate through the baghouse was set to maintain a filtration velocity of 2.0 ft/min during all 4 tests. However, for various reasons the flow rate sometimes varied by as much as 25% from the set point. The parenthesized numbers in the table are calculated estimates of the pressure drops at a filtration velocity of 2.0 ft/min. During the first test, LMF4-B, woven fiberglass bags, 3x1 twill weave, 13.5 oz/yd² with a 10% Teflon coating were installed in the baghouse. The primary reason for using these bags during the initial test was to avoid the possibility of damaging the more expensive Gore-Tex bags during the baghouse shakedown. We also wanted to confirm the bench scale test results indicating that woven fiberglass bags would not work well. As expected, the pressure drop across the bags increased rapidly as the test progressed. After about 3.5 hours the flow rate through the baghouse had to be reduced in order to maintain the draft in the furnace.

TABLE 4. SUMMARY OF 1985 BAGHOUSE TESTING RESULTS

Test: LMF4-B

Cycle Number	1	2	3
Residual ΔP	0.8(0.9)	1.6(3.0)	4.8(5.6)
30 Minute ΔP	3.4(4.3)	6.4(6.6)	9.1(15.7)
Maximum ΔP	5.5(6.0)	9.5(18.4)	9.1(15.9)
Cycle Duration	1:47	1:47	0:52

Test: LMF4-C

Cycle Number	1	2
Residual ΔP	0.3(0.5)	0.9(1.1)
30 Minute ΔP	3.3(3.8)	3.3(3.4)
Maximum ΔP	7.4(10.7)	8.5(7.5)
Cycle Duration	1:30	0:58

Test: LMF4-D

Cycle Number	1	2	3	4	5	6
Residual ΔP	0.3(1.2)	1.2(2.7)	2.7(2.6)	1.8(1.9)	1.8(2.0)	1.8(3.2)
30 Minute ΔP	3.2(4.9)	5.7(5.2)	6.2(6.2)	5.5(6.0)	4.5(6.3)	3.9(8.2)
Maximum ΔP	3.2(5.2)	8.0(8.0)	8.8(6.6)	5.8(9.2)	5.6(9.8)	3.9(8.2)
Cycle Duration	0:32	9:55	0:47	0:49	0:59	0:30

Test: LMF4-E

Cycle Number	1	2	3	4	5	6	7	8	9
Residual ΔP	1.1(1.0)	3.9(3.4)	1.9(2.0)	4.8(3.7)	4.4(3.0)	4.6(3.0)	2.0(1.5)	5.2(6.8)	5.1(5.6)
30 Minute ΔP	5.2(4.3)	8.7(7.8)	6.9(7.0)	8.7(9.1)	8.6(9.0)	10.4(8.5)	5.2(4.5)	-	-
Maximum ΔP	6.9(7.4)	9.5(10.7)	7.4(9.5)	10.9(8.8)	9.7(11.9)	11.1(11.3)	9.0(9.3)	8.8(11.3)	7.3(10.6)
Cycle Duration	0:57	0:54	0:45	9:42	9:52	0:48	1:33	0:27	0:18

(Pressure Drop Unit inches of water)

During post-test inspection it was found that a considerable amount of dust remained on the bags. Average weight gain per bag during the test was 10.4 lb.

For the remaining tests - LMF4-C,D and E - 9.8 oz/yd² woven fiberglass bags with a Gore-Tex membrane laminated on the filtration surface were installed in the baghouse. During these tests it was found that residual pressure drops could be maintained at an acceptable level with the reverse gas cleaning system. However, the rise of the pressure drop immediately after cleaning was very rapid, limiting the filtration time to less than 1 hour per cycle. During LMF4-E sonic horns were used along with the reverse gas at the start of the seventh cycle. Figure 9 is a plot of pressure drop vs. time for the third through seventh cycles during LMF4-E. The characteristic rapid rise in pressure drop after cleaning when only reverse gas was used can be seen in the first 4 cycles on the plot. During the cycle when sonic horns were used the pressure rise after cleaning was much more gradual, requiring about twice as long to return to the former maximum value. This behavior indicates that when reverse gas cleaning alone was used some areas of the dust cake were not being removed, so that when the flow was restarted after cleaning the dust concentrated on the clean areas, accounting for the rapid rise in pressure drop. Using the sonic horns in addition to the reverse gas apparently cleaned the entire surface.

At the start of cycles 8 and 9 of LMF4-E reverse gas cleaning alone was used and was very ineffective. The probable reason for this was an upstream cooling water leak that may have made the dust sticky. After tests LMF4-C and LMF4-D it was found that the bags had gained no weight. After LMF4-E the bags had gained an average of 2.7 lbs each. Collection efficiency for both the woven fiberglass bags and the Gore-Tex bags was about 99.0%, marginally in compliance with the NSPS limit.

In summary, it was found in initial testing that standard woven fiberglass bags will probably not be suitable for MHD dust collection. Results for Gore-Tex bags are encouraging, particularly when sonic horns are used to augment reverse gas cleaning. Long-term testing, planned in 1986, will be required to verify these results.

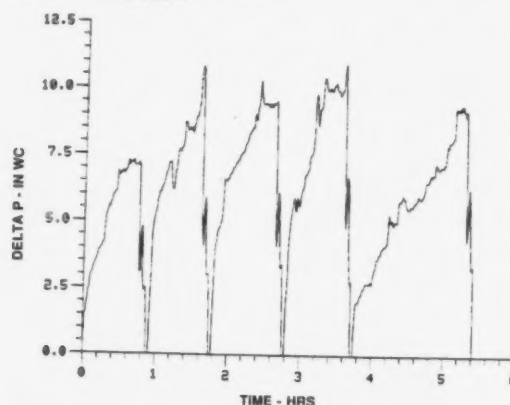


Figure 9. Baghouse ΔP Across Bags Vs. Time During Test LMF4-E

ELECTROSTATIC PRECIPITATOR DESIGN STUDY

The dust properties that were identified as likely to cause electrostatic precipitator operating problems are the adhesive nature of the dust and the small particle size. In precipitators collecting dust from the low odor kraft recovery process, the tendency of the particles to adhere along with the low bulk density causes formation of thick dust layers on both discharge and collection electrodes (4). High intensity rapping is required to remove these deposits and the high rapping force can cause a significant amount of previously collected dust to be reentrained in the gas stream. Problems with tenaciously held dust deposits have also been encountered in MHD pilot testing in Japan. Reentrainment can also occur by flue gas sweepage through the hoppers due to the low density of the collected dust.

A second problem is caused by the effect of submicron particles on the electrical conditions in the precipitator. When there is a large number of submicron particles in the gas stream the charge on the particles, referred to as space charge, can reduce the intensity of the electric field near the discharge electrodes and thus reduce the current flow. This reduces the particle charging rate and collection efficiency is degraded. A related problem, referred to as ion quench, occurs when the number of dust particles approaches that of the available charging ions. This also reduces the charging rate. Problems with excessive space charge do not seem to have been particularly common in kraft recovery precipitators. However, the concentration of submicron particles in MHD dust is probably higher than in kraft recovery dust so a likelihood of space charge problems exists.

The basic electrostatic precipitator design is determined by the type of discharge electrodes, the method of rapping, number of separately energized fields, plate area per unit of gas volume treated or specific collection area (SCA), and rating of power supplies. The two basic precipitator designs in common use by utilities are the "American" design and the "European" design. The American design is characterized by weighted wire discharge electrodes and uses pneumatically or solenoid driven gravity impact rappers. The European design has discharge electrodes attached to a rigid frame and uses mechanically driven hammer type rappers. In the present case, since it is likely that a high rapping intensity will be required on the discharge electrodes, the rigid frame type of electrode was selected. It was also considered important to be able to change the rapping intensity easily, so as to minimize rapping reentrainment. Since changing the intensity of hammer type rappers requires changing the size of the hammer, while gravity impact rappers can be adjusted by changing control settings, solenoid operated gravity impact rappers were selected. The resulting design is referred to as a "Hybrid" design since it has features of both basic designs.

Four separately energized fields were specified. This is common practice for precipitators designed for over 99% efficiency (7). Having four fields allows for optimization of electrical conditions, which vary from front to back of the precipitator due to the changing particle concentration.

Determining the required SCA for a new process is difficult because the

physics of the precipitation process are so complicated. Unless there is considerable operating experience for precipitators on similar processes, pilot studies are usually required. In our design study we looked at operating experience of ESP's on kraft recovery boilers, and previous pilot studies on MHD (or simulated MHD) dust. We also used an electrostatic precipitation computer model developed by Southern Research Institute to predict the required SCA to reach the design efficiency of 99.6% (8), (9). In kraft recovery operations the 99.6% efficiency level has been met in most cases by designing ESP's for SCA's of 450 to 500 ft²/1000 cfm. Three previous pilot tests on MHD or simulated MHD dust have been conducted, including a bench scale test using a tube type precipitator to collect dust produced by a fuel oil-ash slurry; a test on a full scale precipitator where K₂CO₃ was added to the coal fired in a pulverized coal boiler; and tests at a pilot scale oil-fired MHD research facility in Japan. Results of these tests varied so widely that they are of little use for sizing the CFFF precipitator. By using the Deutsch equation to extrapolate the results to 99.6% efficiency, a range of SCA's from about 200 to over 800 was obtained. In the computer model study, dust properties of samples collected during testing at the CFFF along with geometric parameters consistent with standard precipitator design were used as inputs to the model. A parametric study was undertaken to determine the effects of varying dust size, plate-to-plate spacing and electrical conditions on collection efficiency. The main drawback of the model is the lack of a means for accurately determining the voltage level at which sparking will occur. (There is an empirical routine for determining sparking voltage, but it was developed only for coal fly ash applications). This makes determination of operating voltage somewhat arbitrary. An interesting result of the study is the potential effect of excessive space charge and ion quench on collection efficiency. Figure 10 shows a plot of collection efficiency vs. SCA for a precipitator with "normal" electrical conditions and for a precipitator with a reduced charging rate due to excessive space charge and ion quench. In the plot it can be seen that the quench effect is reduced as the SCA is increased. This is due to more particles being removed in the upstream part of the precipitator, thus reducing the quench effect in the downstream section. The study predicted required SCA's for a 99.6% collection efficiency ranging from 310 for a precipitator with normal electrical conditions to 450 for one with space charge problems. Based on the above considerations, a design point SCA of 475 was selected for the LMF precipitator, with possible test conditions ranging from 350 to 750 to cover the range of practical interest.

Rating of the power supplies was somewhat arbitrary, since there was no way of predicting electrical operating conditions accurately. It was thought that the possible effects of space charge, which would tend to increase the corona starting voltage, combined with the low resistivity of the dust, which would tend to increase the sparking voltage, might allow the precipitator to operate at higher than normal voltages. Therefore the transformer-rectifier sets were rated at 75 kV_{av}. A current rating of 120 μ A/ft², about twice the maximum predicted by the computer code, was specified to allow for sparking. SCR type controls were specified.

For a SCA of 350 ft²/1000 cfm at a flow rate of 8500 cfm, collection plate area was set at 3000 ft². Maximum bulk gas velocity was limited to

4.0 ft/sec to reduce the likelihood of reentrainment.

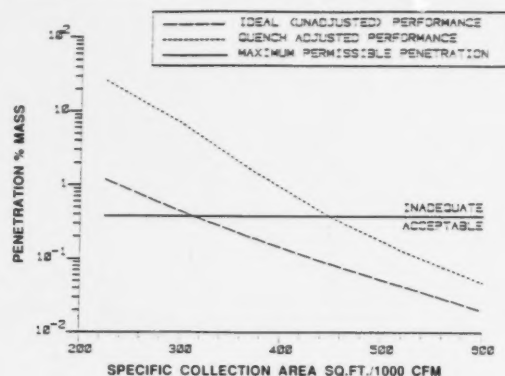


Figure 10. Collection Efficiencies Predicted By Computer Model

The electrostatic precipitator is shown in Figure 11. The shell is constructed from 3/16 inch carbon steel plate with 4 inches of external insulation. The collecting plates are 10 feet high by 9 feet long and are constructed from 16 ga carbon steel sheet. There are four 12 inch wide gas passages. The discharge electrodes are a combination of 1/4" square and 1/4" diameter round wires which are hung from the top and guided at the bottom. Each section has 9 wires, 4 square and 5 round. Rappers are located on the roof at the front and back of each field. An electric vibrator is attached to the discharge electrode support frame in each field. There are two perforated flow distribution screens at the inlet and one at the exit. A single trough hopper equipped with a screw conveyor runs the entire length of the precipitator. It has a separate discharge for each field and is baffled between fields to prevent gas sneaking.

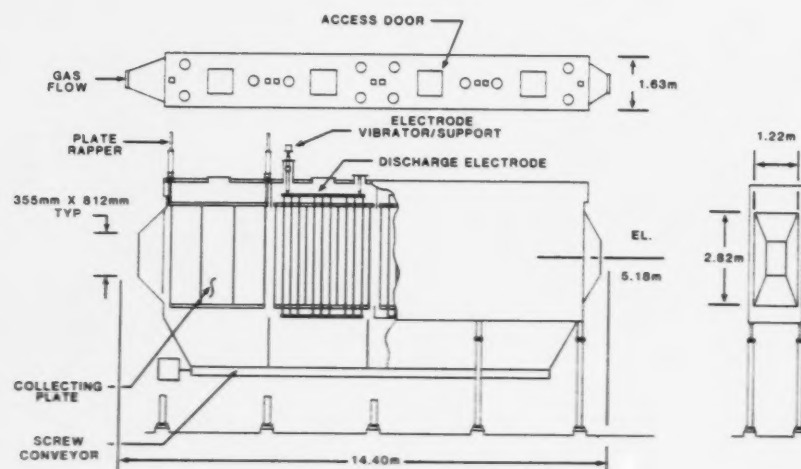


Figure 11. LMF Electrostatic Precipitator

INITIAL ELECTROSTATIC PRECIPITATOR TEST RESULTS

The ESP was operated during two tests in the fall of 1985 for a total of about 14 hours. Gas flow was regulated to maintain a SCA of 475 during both tests. During the first test, LMF4-D, the precipitator did not operate correctly, as all four electrical sections sparked at very low voltage. During post-test inspection it was found that the support insulators for the discharge electrodes were wet and coated with dust. It was determined that the low operating voltage was caused by arcing across the insulators. Dust collection during this test was negligible.

It was determined that the moisture on the support electrodes condensed from flue gas because the insulators were not hot enough at the start of the test. To prevent this problem from recurring during the second test, LMF4-E, the insulator enclosures on the precipitator roof were insulated and the precipitator was preheated by firing a fuel oil burner into one of the hopper level access doors.

Voltage and current in the ESP fields at the start of LMF4-E are shown in Figures 12 and 13. For comparison, a clean-plate air load voltage-current curve obtained before the test is shown in Figure 14. During the early part of the test, voltages of over 50 kV_{av} were reached in fields 1 and 2 and there was a significant current flow in fields 3 and 4. Although there was no quantitative measurement of collection efficiency, observation of the gas stream at the exit of the precipitator indicated that collection efficiency was good. After about 40 minutes the voltage limit on the first, second and third fields was set to limit the maximum voltage to 75 kV. This resulted from a misunderstanding between us and the T-R set vendor (we thought the secondary voltage meters indicated average kV, while they actually indicate peak kV). This is the reason for the drop in voltage shown in Figure 12 at 0.65 hours. At about 0.75 hours field 1 shorted out due to rain water leaking through an incorrectly installed gasket on the insulator enclosure roof. At about 0.90 hours the same problem occurred in field 2. Current flow in fields 3 and 4 decreased rapidly during the first hour of the test and was negligible after about 1.5 hours. The reduced current flow was probably caused by a coating of dust up to 1 inch thick on the discharge electrode wires that could not be removed by the vibrators. The test was interrupted after about 2 hours to fix the leaks and when it was resumed all four fields returned to their former operating points of about 75 kV. There was still no significant current flow in any of the fields. Collection efficiency, as measured by mass train dust loading tests at the precipitator inlet and exit, was about 87% during the latter part of the test.

In spite of the series of problems encountered during the test, the results were not entirely discouraging. The problem of excessive space charge, while it may account for the low current flow in fields 1 and 2, did not seem to extend to fields 3 and 4 since there was initially a significant current flow in those fields. The inability of the vibrators to remove dust from the discharge wires is believed to be due to a Teflon insulator used to connect the vibrator to the wire frame. The Teflon insulator acts as a damper and reduces the intensity of the vibration transmitted to the wires.

These insulators will be replaced by more rigid ceramic insulators prior to long term tests planned in 1986.

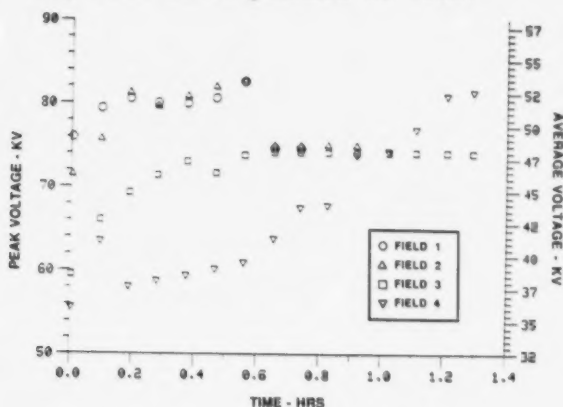


Figure 12. ESP Secondary Voltages During Test LMF4-E

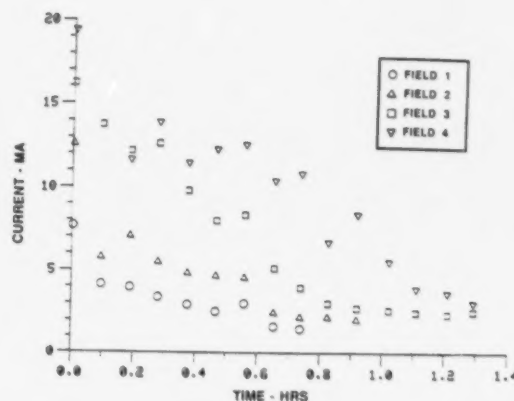


Figure 13. ESP Secondary Currents During Test LMF4-E

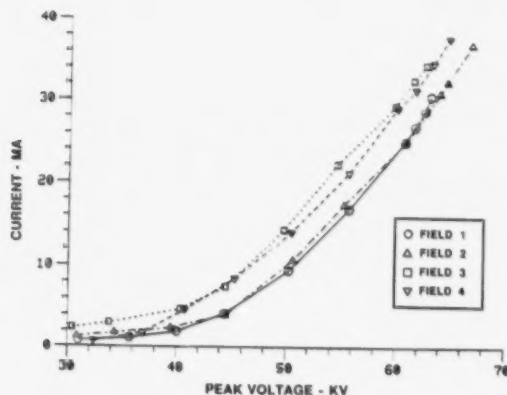


Figure 14. ESP Air Load Voltage-Current Curve

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GRANULAR BED FILTER COMPONENT DESIGN
FOR HIGH TEMPERATURE AND HIGH PRESSURE

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ABSTRACT

A Granular Bed Filter is being fabricated by Combustion Power Company for operation on a pilot-plant-scale PFBC at New York University, Westbury, NY. As part of DOE's program to develop coal-fired pressurized fluidized-bed combustors (PFBC) for use in combined-cycle, power generating systems, a hot gas cleanup train must be used on the pressurized gases before entering the gas turbine to ensure removal of a major portion of particulate. The Granular Bed Filter (GBF) has shown considerable promise to date and has been chosen for further investigation by DOE.

The objectives of this program are 1) to augment information on the feasibility of a GBF for controlling particulate in a PFBC power generating system, 2) to assess the design and life of critical components in the filter system, and 3) to estimate costs of a commercial-scale design based on tested concepts.

Life-critical component testing at Combustion Power Company is in progress. The tested components are essentially the same items to be installed at New York University (NYU) with the exception of the filter vessel. Results from these tests will be combined with results from upcoming NYU tests to prepare a design and cost estimate for a commercial-scale GBF.

Findings are expected to demonstrate that the GBF concept will be an operationally and economically feasible means to remove particulate from PFBC off gas and will meet turbine tolerance limits and Federal pollution control standards.

This paper has been reviewed in accordance with the U.S. Environmental Protection Agency's peer and administration review policies and approved for presentation and publication.

INTRODUCTION

Efforts to operate coal-fired gas turbines have been conducted in recent years. These attempts, like the one at Combustion Power Company in the early 1970's, have been plagued by turbine problems due to ash-laden combustion gases. The U. S. Department of Energy is currently sponsoring programs to develop coal-fired, pressurized fluidized-bed combustors (PFBC's) to be used in combined-cycle power generating systems. These systems utilize a gas turbine driven by pressurized combustion gases to generate a portion of the electricity. It is recognized that a hot gas cleanup train must be used before the gas turbine to remove the major portion of the particulate. This is necessary to prevent erosion of turbine materials and deposition of particles within the turbine. The Granular Bed Filter (GBF) has shown considerable promise to date and has been chosen for further investigation by the U. S. Department of Energy.

Commercial application of the equipment and principles being evaluated in this effort will result in a power plant configured much like that shown in Figure 1. This figure shows four GBF modules applied to a 600 MWe power plant.

In the early 1970's, Combustion Power Company (CPC) began experimentation with a method of cleaning exhaust gases or process streams of particulates using granular media. The media was in the shape of 2 mm alumina spheres. This experimentation was in conjunction with PFBC development for power production utilizing a gas turbine. The gas cleaning process used is reminiscent of filtering suspended solids from wastewater through a sand bed.

After a few early fixed-bed tests and a short flirtation with the use of puff-back to clean these beds, techniques were implemented to use a continuously moving bed filtration device. At this time, the work at CPC took two paths; low to medium temperature gas cleanup and high temperature gas cleanup.

HOT GAS CLEANUP

Initial hot gas cleanup testing utilized the same technology being used for low to medium temperature applications, that is the 2 mm granular media was contained in the space between two concentric louvered screens. Figure 2 shows an early design of CPC's high temperature granular bed filter (GBF). Performance of this filter was quite good but when tested on the actual exhaust of a 2000 acfm coal/dolomite fired fluidized-bed combustor, re-occurring plugging of the front (inlet) face resulted (Guillory, 1980). A number of techniques were tried to eliminate this problem. The most successful of these was the use of prearranged media spillage through the screens across the front face. This technique was first used by the commercial version of the filter where it has been patented. Even using this device, front face pluggage eventually resulted when the filter was operated on the fluidized-bed pilot system at temperatures over 1400°F.

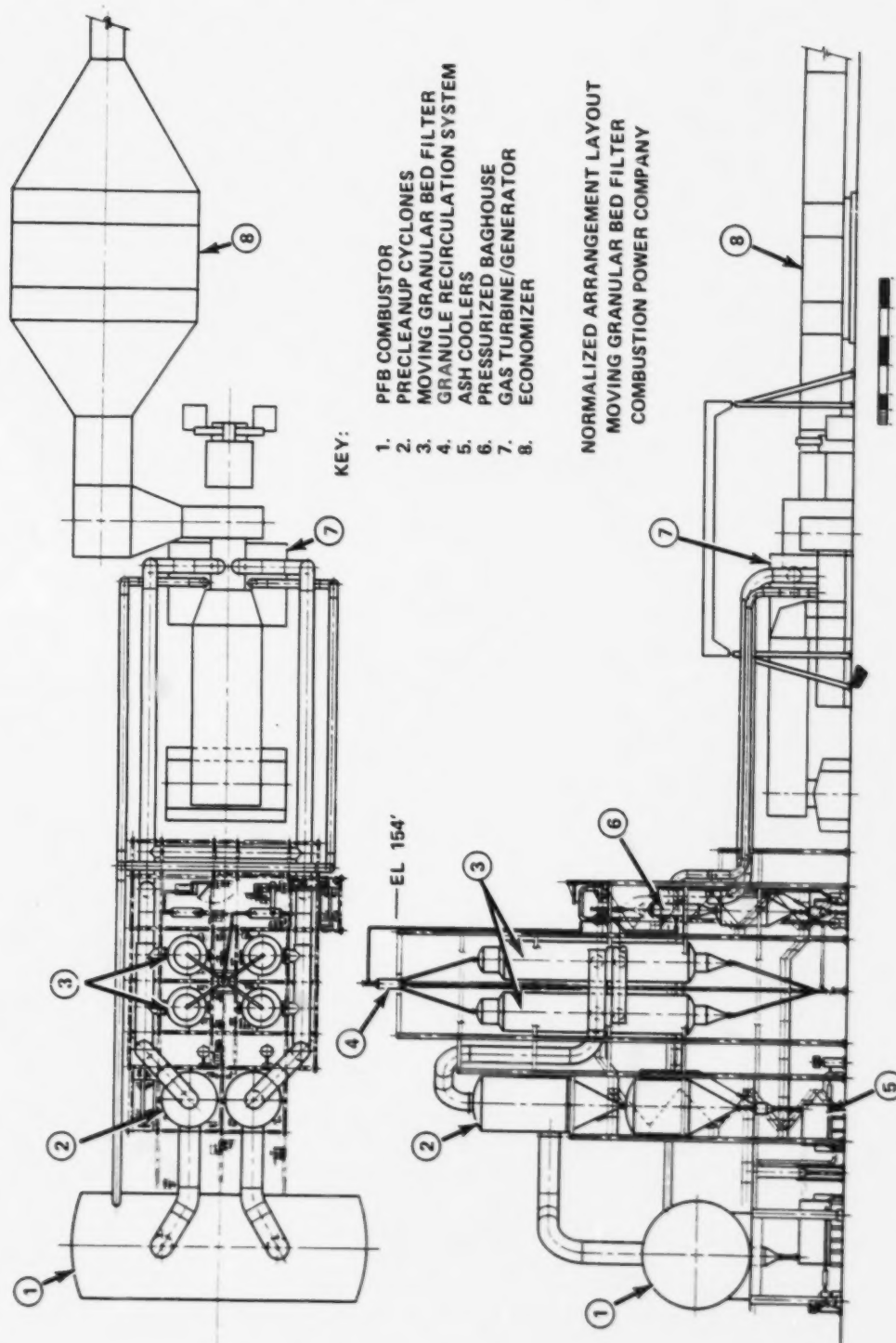


Figure 1 Granular Bed Filter in a Combined-Cycle Power Plant Train

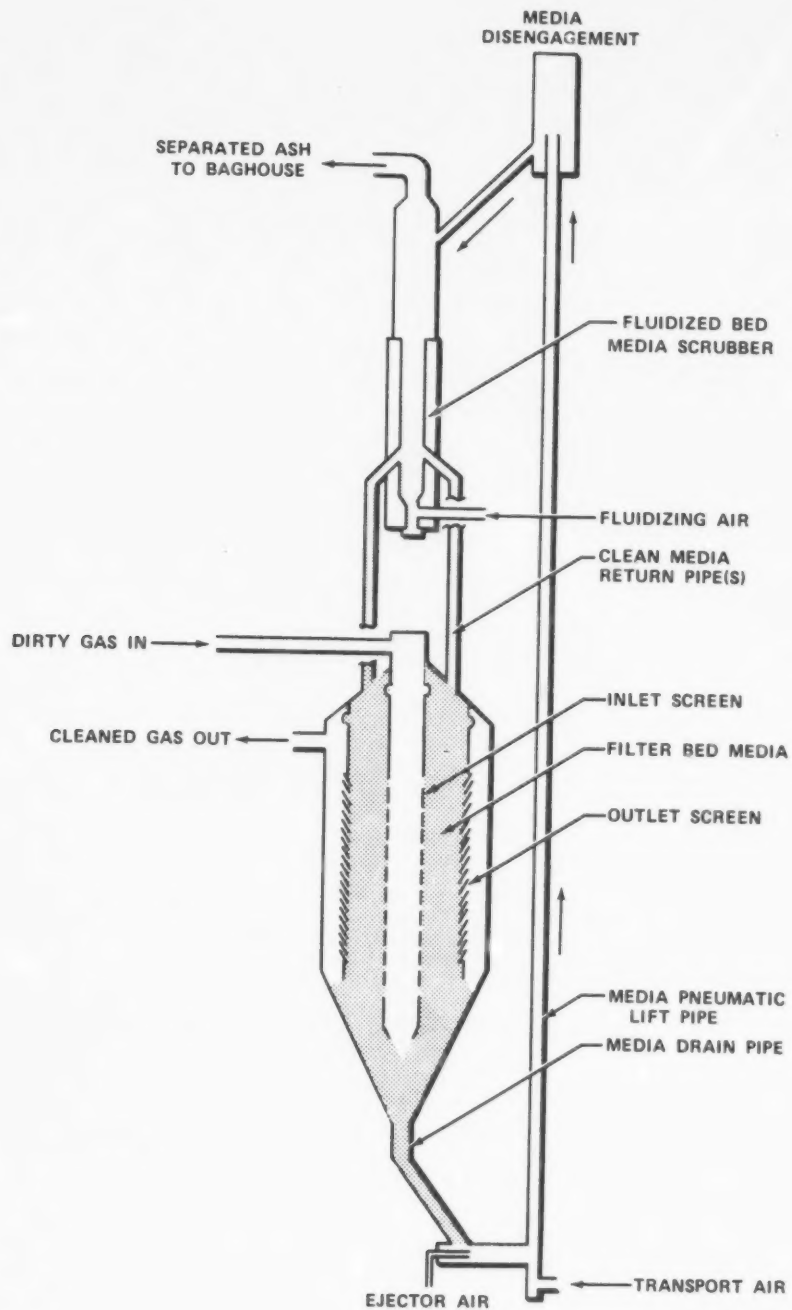


Figure 2 Screened Cross-Flow Granular Bed Filter

During the testing of the "screened" filter, it was observed that when the front face was plugged, pressure drop across the filter rose only nominally (to ~ 25 in. H_2O) and leveled off. Simultaneously, filtration performance increased. These observations led to a totally new approach to implementing hot gas cleanup using a GBF device. Figure 3 contains a schematic view of the present "screenless" configuration (Guillory, 1983). This new filter is no longer a cross-flow device, but combines features of parallel-flow, cross-flow and counterflow with the latter at the gas outlet.

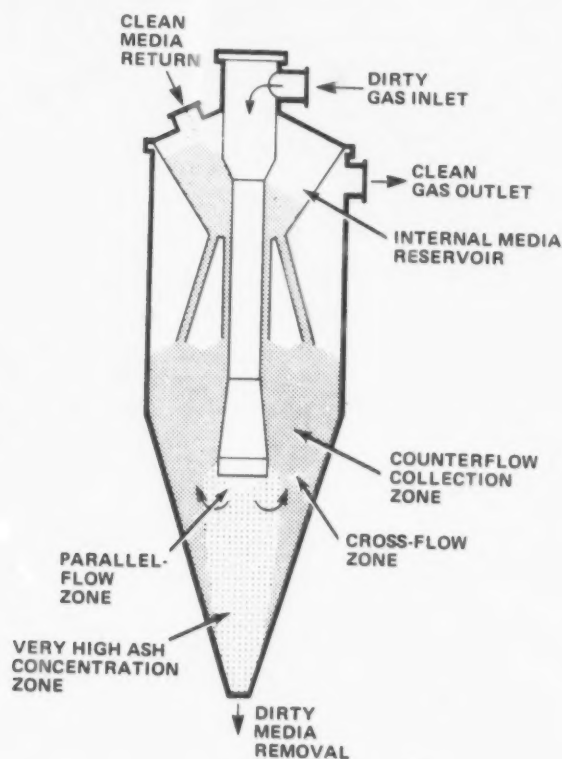


Figure 3 Screenless Counter-Flow Granular Bed Filter

Looking at these various zones, one can see the benefits attained by each:

- Parallel-flow zone: At the filter bed inlet, 90% of the solids are removed by a small portion of the media. The result is a semi-cake which promotes even more ash collection. The rapidly moving media in this region and the shear layers near the boundaries promote tumbling and mixing of the media. This action keeps the ash from sticking either to itself or to the media. Thus, a large amount of re-entrainment of the ash occurs in this area of high media and gas motion.
- Cross-flow zone: The adjacent cross-flow zone recaptures most of this ash carrying it back into the high velocity core where the ash is literally drawn down and out of the filter vessel with the media.
- Counterflow zone: Here the final ash collection-polishing takes place. The GBF is a true counterflow device in that exhaust gases exit through the cleanest media. In fact, one can always design with additional layers of media at the exit to increase cleanup effectiveness. This knowledge can even be used, theoretically, to size filter elements for larger (3 mm) alumina media.

While the original filter configuration had a pressure drop in excess of 25 in. H₂O, it provided very high performance. Nearly three thousand hours of high temperature operations were accumulated with the "screenless" filter. The final two thousand hours were with the latest design configuration, and one thousand hours of testing concluded the development program. This final test was conducted in segments of 100, 200, 300 and 400 hours. After each segment, the filter was shutdown, test data evaluated and weeks later testing was resumed. No extraordinary work was ever required to restart the filter, and no cleaning or modification was ever made to the filter. At the conclusion of the one thousand hours, the system was drained and fully inspected. There were no signs of pluggage, agglomerations, excessive wear, or any extraordinary problems. During the tests, all operations performed smoothly. Special tests which were performed included media stoppage and routine restart, and planned loading upsets of up to 5 gr/acf to simulate PFB upset conditions; the filter took it all in stride (Guillory, 1980).

Figure 4 displays the efficiency data from this test series in graphical form. It was found that the data was able to be correlated to the normalized parameter shown. With the proper selection of design parameters, New Source Performance Standards can be met while providing a gas turbine with economic blade life.

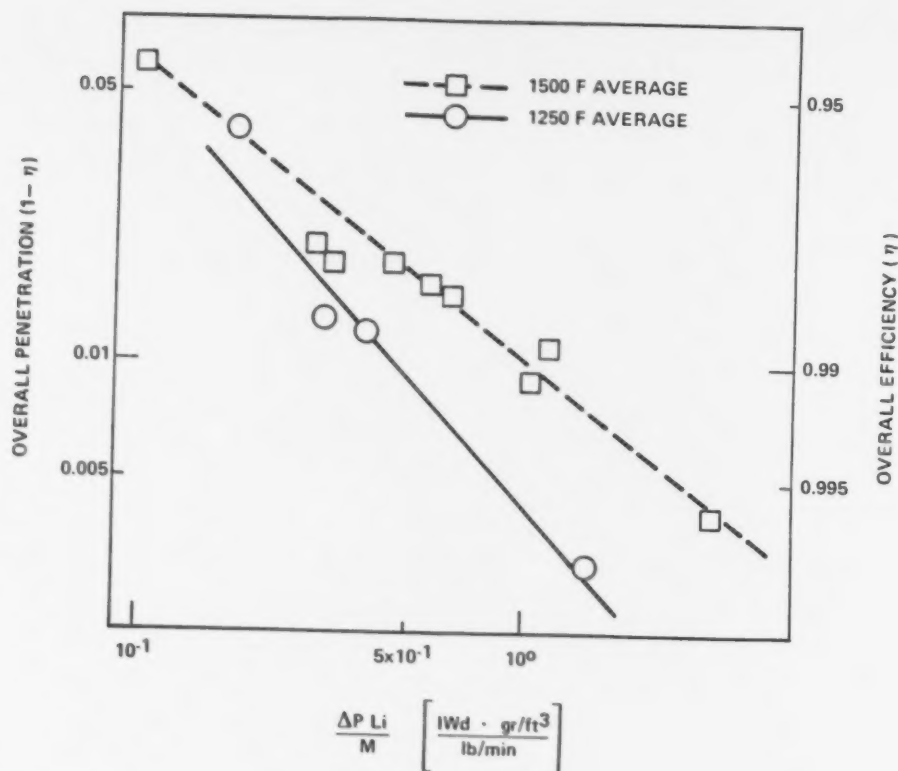


Figure 4 Particulate Collection
Performance 24-In.
Counterflow (Screenless)
Granular Bed Filter

PLANNED HIGH-TEMPERATURE, HIGH-PRESSURE DESIGN AND TESTING

The purpose of this development effort is to obtain information on the operational and economic feasibility of the moving granular bed filter in a PFBC power generating system. Information generated will consist of (1) operational data carried out on the pilot-plant scale, (2) experimental data generated on critical GBF system components, and (3) an update of the commercial-scale design based upon the experimental data.

To meet the stated purpose, the contract is divided into three tasks as follows:

TASK 1: A pilot plant test sequence at New York University utilizing actual PFBC exhaust.

- The objective of the test sequence is to successfully remove particles to the required levels over a 200-hour test period without operational or component failure. Testing includes filtration experiments with 2 mm and 3 mm alumina media spheres.

TASK 2: A critical component test sequence to be carried out at Combustion Power Company in Menlo Park, CA.

- The objective of this effort is to design and test critical components necessary for GBF operation at high temperature and high pressure. These critical components are the media circulation and ash handling systems. 500 hours of component and parametric tests are planned.

TASK 3: An update of a commercial scale concept for a PFBC power station. This concept was previously developed based on testing at high temperatures and low pressures.

- The objective is to estimate capital and operating costs of a cleanup train, based on the GBF concept and design principles.

Designs for the test facilities at CPC and NYU are complete. Equipment for critical component testing at CPC is installed and testing is underway.

GBF EQUIPMENT FOR NEW YORK UNIVERSITY TESTING

The GBF gas cleanup system consists of three main subsystems:

- The granular bed filter element in which the particulates are removed from the incoming gas stream.
- The media circulation system which continually withdraws dirty media from the apex of the element cone and returns cleaned media to the top of the element.
- The ash handling system which transports the particulate separated from the media to the ash storage bin.

The subsystems are shown interconnected on Figure 5. PFBC flue gas flows through the moving granular bed and gives up particulate. 2 mm or 3 mm filter media, laden with ash, flows downward towards the apex of the filter cone to be collected for cleaning and recirculation.

The filter media will be recirculated at a rate of 15-40 lb/min using nominally 11 lb/min of lift gas by a 2-inch pneumatic lift pipe. The transport gas will be flue gas that is being recirculated through the closed transport gas system. As the particulate-laden filter media is transported vertically through the lift-pipe, the media is cleaned. At the top of the

lift-pipe, a de-entrainment vessel separates the cleaned filter media from the now particulate-laden transport gas. The cleaned filter media then flows by gravity to a media reservoir located above the filter pressure vessel.

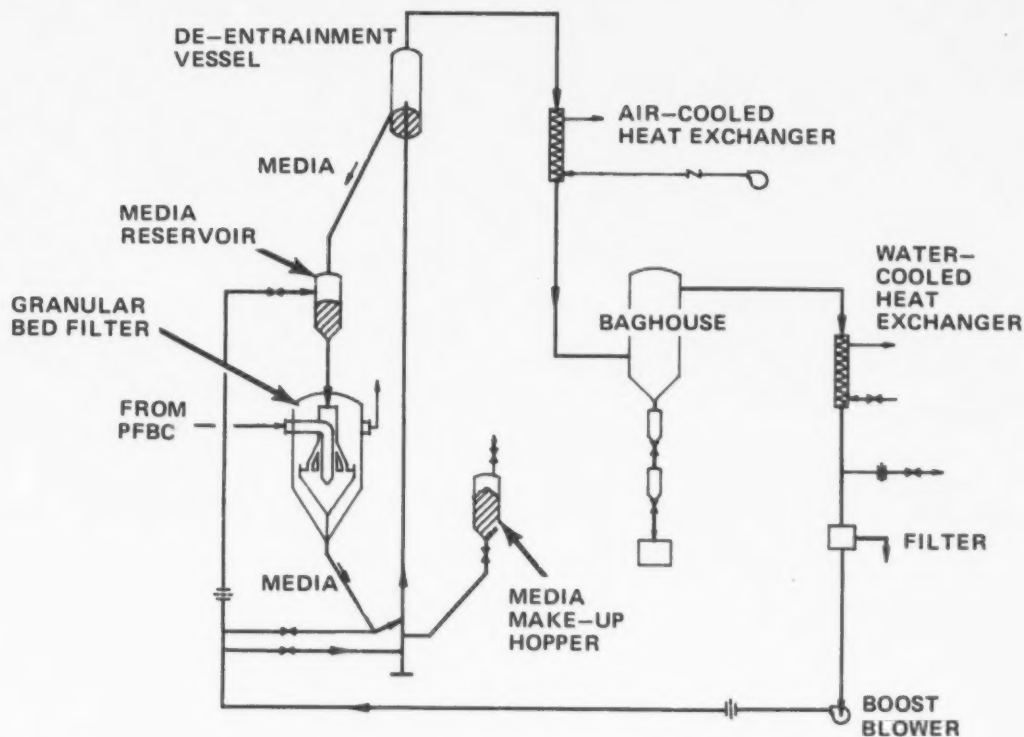


Figure 5 Schematic for Subpilot Testing at NYU

The hot, particulate-laden transport gas leaving the de-entrainment vessel will be cooled and sent to a pressurized baghouse for removal. Cooling will be achieved first by radiation in a 40 ft. long uninsulated, stainless steel pipe, then by convection in an air-cooled heat exchanger. Temperature entering the baghouse will be nominally 500°F. The air-cooled heat exchanger is designed for moderately high velocity inside the tubes to minimize fouling. In operation, the tube wall temperature is maintained above the dewpoint to prevent moisture from condensing on the tube walls and attracting ash.

The collected dust from the baghouse will be transported to a disposal bin by a double lock-hopper arrangement. The cleaned transport gas will be recompressed and circulated back to the media transport system.

For successful operation, the media circulation and ash handling systems must exhibit the following key features:

- The media returns clean (dust free) into the filter.
- Plug flow of media in the seal legs restricts gas flow from the filter cavity into the circulation system.
- Media is cleaned while hot, and returned to the filter at the lowest possible heat loss.
- The lift gas transports the separated particulates to the ash removal equipment.
- The pressure in the circulation system is controlled, if necessary, to assist ash removal from the filter.

The system is designed so the pressure at the bottom of the lift-pipe can be made equal to or slightly less than the GBF outlet pressure. This control is achieved by bleeding circulation gas to atmosphere just downstream of the water-cooled heat exchanger. The advantage is that the filter element at the apex and the lower seal leg operates at essentially zero reverse air flow. This condition is favorable for reliable operation, in that gas leakage upwards from the media valve area does not retard ash from entering the seal leg. At the top of the GBF element, another seal leg essentially seals process gas from the media circulation system. The pressure drop across this section can be made equal to the lift-pipe pressure drop by adjusting the amount of circulation gases bled to the atmosphere. If, on the other hand, no gas is bled to the atmosphere the pressure drop due to transporting media up the lift-pipe (1-2 psi) will cause a small amount of circulation gas to bypass the lift-pipe, and flow through the filter element to the de-entrainment vessel. The velocity of these leakage gases, however small it is, may cause ash to dislodge from the media and concentrate at the apex of the filter cone. The design of the media circulation system allows flexibility to counteract this phenomena if it occurs.

The arrangement of the filter element within the pressure vessel is shown on Figure 6. In comparing this configuration with the filter element that underwent 1000 hours of low pressure, high temperature testing (Figure 3), one recognizes that the critical relationship between the physical components within the filtering regime has remained unchanged. The dimensioning between the inlet gas duct near its outlet into the filter bed and the filter walls and media distribution pipes remains essentially unchanged. The intent is to preserve the gas flow distribution through the parallel-flow, cross-flow, and counterflow collection zones. Media distribution to the filter body has been altered significantly as can be seen by comparing Figures 3 and 6. Moving the media reservoir out of the filter has advantages in the commercial-scale design.

Relocating the media reservoir to above the filter element saves valuable space in a pressure containment vessel. Furthermore, the media feed pipe from the remote reservoir creates a low-leakage seal between the filter

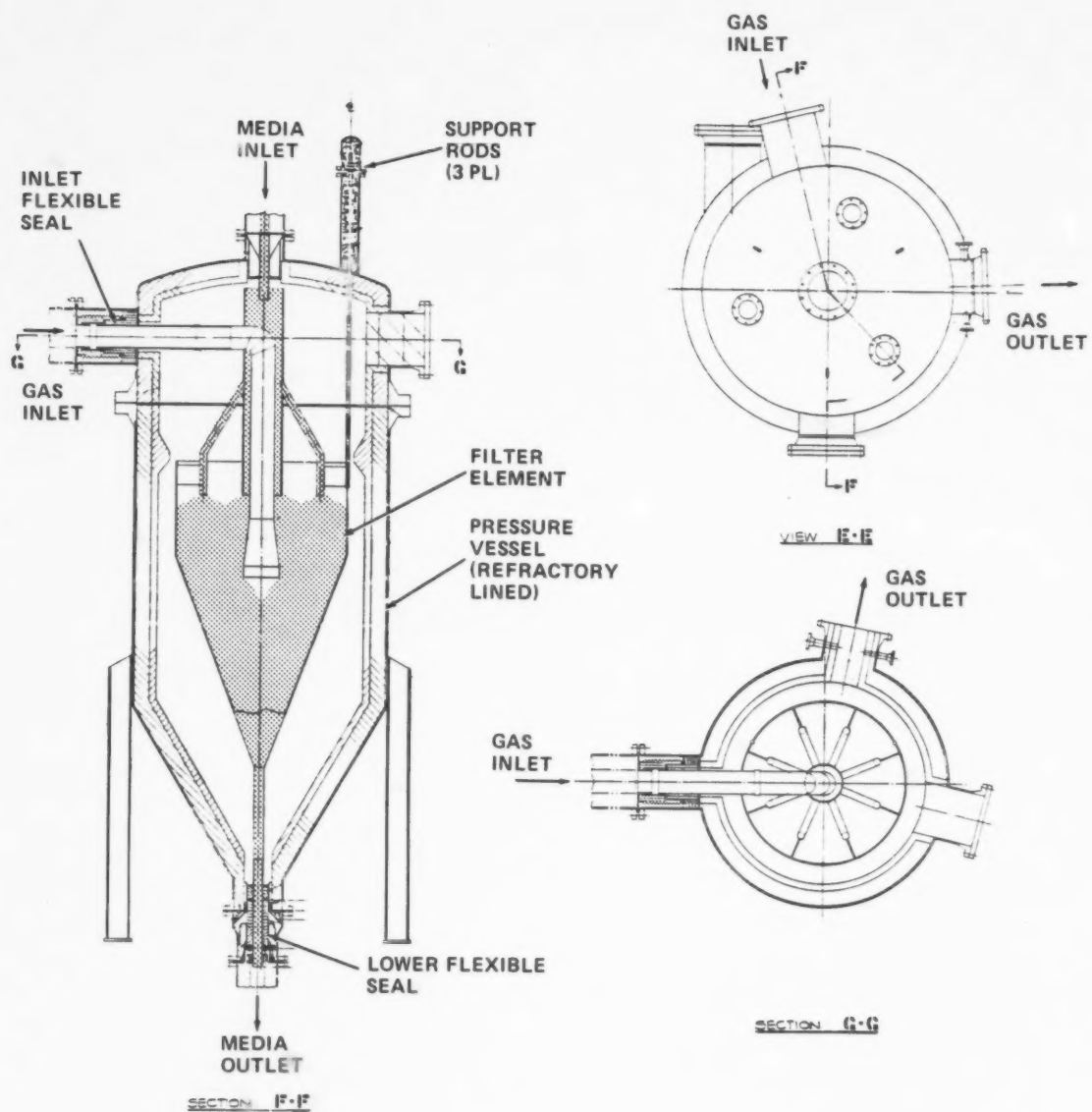


Figure 6 Granular Bed Filter Installation

element and the media circulation system. Minor pressure differences between these two zones can exist without upsetting media flow or particulate capture.

There are, although, certain risks associated with altering the media flow paths. CPC's experience with the 2 mm spherical ceramic media has been quite extensive. It includes thousands of hours of observations and data collection. In each case of flow alteration we can point to similar cases of successful media movement.

At the filter inlet, a flexible seal absorbs small horizontal and vertical movements from thermal expansion. The basis of the seal is a metal bellows that can absorb the thermal movements and maintain a positive seal against inlet gas bypass of the filter bed. At the media outlet, another flexible seal absorbs considerable (about 3½") of vertical thermal expansion. This metal bellows is also insulated from the 1600°F environment so that common, stainless steel materials can be used in fabrication.

To support the filter element within the pressure vessel, a simple design approach was used. This approach is patterned after common industrial approaches to supporting metal components inside refractory-lined furnaces. The support rods are carefully designed to allow a temperature gradient along the rod which is steep enough not to allow overheated support points. Heat loss due to this configuration is quite modest, only a few percent of the total. The limiting factor is the small cross section of the rods.

EQUIPMENT FOR CPC TESTING

Figure 7 shows a schematic of the system to be developed and tested at CPC prior to the NYU test sequence. This system includes the actual media circulation and ash handling components to be utilized at NYU. The equipment, operation and flexibility of the system is much like that described for installation at NYU with the following exceptions:

- Since the GBF and pressure vessel technology is developed further than the remainder of the system, this unit will be replaced with a small surge hopper that partially simulates the GBF. Arrangements will be made to inject ash into the GBF simulation vessel.
- The heat source to bring the NYU filter system up to temperature is the PFBC. A separate heat source is needed for critical component testing at CPC. The system heaters shown will not be included for NYU operations.

Figure 8 shows an elevation view of both the CPC critical component test system and the NYU pilot-scale filter. These views are unfolded for clarity, but are depicted somewhat to scale, vertically, to give an idea of the physical differences. The 10 foot height difference between the NYU and CPC test facilities saves cost in the support structure at CPC and in the GBF simulation vessel. Operational differences due to this change are not significant.

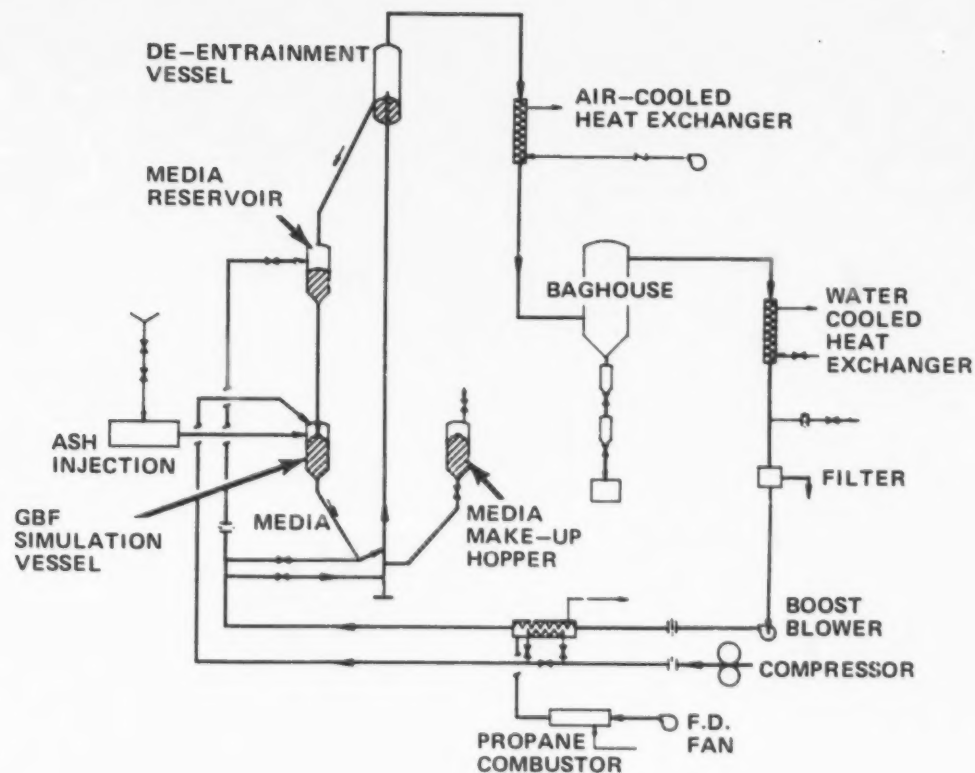


Figure 7 Schematic for Critical Component Testing at CPC

Some of the salient features and basis of the circulation and ash handling system design are:

REFRACTORY LINING:

The lift-pipes and seal-legs were made from numerous short sections of refractory-lined pipe. Lift-pipe segments were 12" standard wall pipe lined with two layers of castable refractory down to a 2" inside diameter (I.D.). The insulation layer is a soft castable block mix. The hard face is a 3" thick layer of abrasion resistant castable. This installation was performed using common materials but special techniques developed for producing refractory-lined pipe. Forms were made from heavy cardboard tubes centered in the pipe with plywood forms. Each section of finished pipe was checked for accurate centering and straightness of the resultant refractory-lined hole. Tools used for checking included bolthole sized tapered pins and a light-gauge plate drilled to the proper bolt circle and containing an alignment hole in the center. The intent was to produce refractory-lined pipe with the I.D. centered within flange bolt circle. With this fact proven, the flange bolt holes could be used with the tapered pins to line up 5' sections of pipe so the I.D.'s at each flange interface lined up. The tolerance

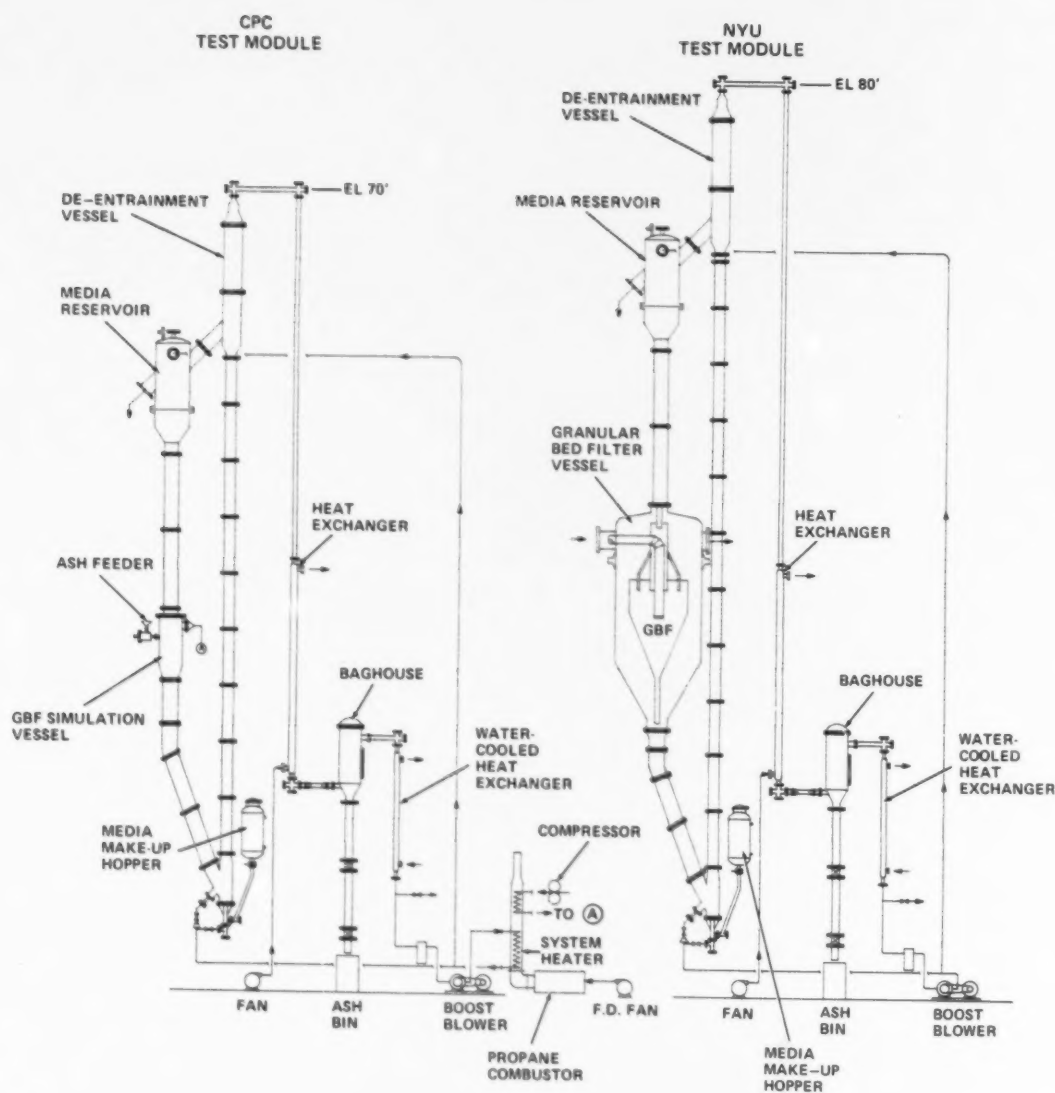


Figure 8 Test Facilities to be Installed
at CPC and NYU

chosen for alignment was $\pm 1/16"$ based on engineering judgment. The refractory installation job proved adequate and no work had to be rejected.

Installation of the piping was completed by lining up match marks on the component flanges and then lining up boltholes with two tapered pins per joint. Proof of proper alignment was achieved by sliding a 1-15/16" diameter wooden dowel past each mating flange. The dowel had an adjustable length PVC

pipe handle for convenience. In the few predictable instances that alignment was not immediate, minor levering at the boltholes accomplished the desired result. The same tapered pins were used on each flange. As bolts are normally 1/8" smaller in diameter than the boltholes, there is some adjustment allowable at the flanges. Seal-legs were made by lining 14" pipe to a 3" I.D. The same principles were utilized.

DE-ENTRAINMENT VESSEL:

The design of the de-entrainment vessel which separates media and ash by velocity is patterned after equipment developed in the high-temperature, low pressure testing at CPC. The 2" I.D. lift-pipe is extended into the bottom of the de-entrainment vessel as a stainless steel pipe. The remainder of the vessel body acts like an air classifier in that the gas velocity is adjusted to allow media to drop out of the gas stream while ash is carried upward. Baffles near the top exit prevent media escape in the event of an unusually high gas velocity.

MEDIA RESERVOIR:

In low pressure GBF testing at CPC, the media reservoir was an integral part of the filter vessel. For high pressure operation, the media reservoir is physically separated from the filter vessel to minimize gas leakage between the filter and circulation system, and to lay groundwork for scaling up to a commercial-size filter.

MEDIA VALVE:

Media circulation experience in previous GBF testing and in commercial filter design (300-500°F and atmospheric pressure) formed the basis for configuration and sizing of the media valve, lift-pipe and seal legs.

HEAT EXCHANGERS AND BAGHOUSE:

Choice, arrangement, and application of the heat exchanger and baghouse comes from utilizing the inputs of qualified vendors to apply this equipment within its applicable range.

ASH UNLOADING:

The Ash unloading valves and sequencing is patterned after similar systems for high pressure cyclones utilized in developmental PFBC exhaust systems.

BOOST BLOWER:

The boost blower choice of a rotary vane compressor operating well within its pressure and temperature range results from a comprehensive study of candidate machines.

SUMMARY

The 500-hour, critical component test sequence at CPC should serve to demonstrate the operation of the media circulation and the ash handling components. Further objectives are to identify the operating envelope for high-temperature, high-pressure media circulation; to reconcile the media transport model with resulting data; to observe the effects of gas leakage through the seal-legs; to determine limitations on ash/media concentrations; and to check for wear and media attrition.

Equipment and operating data from critical component testing at CPC will be used in the 200-hour, NYU test sequence. Tests at NYU will concentrate on assessing the particulate removal efficiency of the GBF in the environment of an actual PFBC exhaust. Parametric tests will be conducted over a pressure range of 7-10 atm at different process gas flows and media circulation rates. A single operating point will be chosen for a long term (100-hour) steady state test.

Based on results from both test sequences, the commercial-scale approach will be updated and cost estimated.

CPC's goal is to demonstrate the GBF concept to be an operationally and economically feasible means to remove particulate from a PFBC off gas. We expect the GBF to meet both turbine tolerance limits and Federal pollution control standards.

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THE INTERACTION OF A GAS WITH SOLID AND LIQUID PARTICLES IN A ROTATING SYSTEM

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ABSTRACT

A set of Navier-Stokes equations is given in a three-dimensional cylindrical coordinate system that, in addition to the inertia and viscous terms, includes centrifugal and Coriolis forces. These equations are then applied to two-phase flow systems, one a gas-solid stream, the other gas (or vapor) - liquid. For both systems the analysis takes gas velocity as an input and predicts particle trajectory as a function of its initial characteristics.

For solids, the analysis also determines the ricochet path of the particle as determined by its coefficient of friction and geometry. The analysis can be applied to particulates in the turbine, compressor, or fan blading in order to determine their impact on the walls, or to particulate elimination by means of centrifugal separation. For the gas-liquid system, the analysis determines the relative velocities and trajectories of the droplets in the field of a rotating gas or vapor and their rate of vaporization as a function of the prevailing flow field and inlet conditions. The computer-obtained results provide the conditions required to avoid collision with the walls. These results can be applied to turbine blades in which the steam contains either moisture or solid particles with a view of minimizing blade erosion.

INTRODUCTION

In its general context, this paper deals with the dynamics of particles in a three-dimensional gaseous flow field. The velocity map of the gas or vapor, whatever its form, is assumed to be given and the trajectories of the particles are obtained in terms of the flow field, particle characteristics, entry conditions, and the geometry of the configuration in which the flow occurs. Of particular interest here is whether or not and under what conditions the particles are likely to impinge on the walls. The flow of solid particles, within the geometry of a centrifuge, is investigated with reference to a fixed-coordinate system and can be considered as occurring in an actual centrifuge, or in the nozzles of a turbine, or any other stationary passage. The flow of liquid

particles is studied in the rotating field of compressor blades and their trajectories are calculated relative to the rotating coordinates fixed in the blades. These results are directly applicable to the blades of any turbomachine. In any case, the conversion of a stationary-coordinate system to that of a rotating one can be accomplished by the simple addition to the circumferential velocity, ωr . The study of liquid particles differs from the solid ones in that during the passage of the droplet through the blade spacing, evaporation of the droplet is taken into account. On the other hand, solid particles ricocheting from the walls was considered, whereas the droplets instead of rebounding would disintegrate on impact.

SOLID-GAS FLOW

EQUATIONS FOR GAS-SOLID PARTICLES

Throughout the paper those the quantities referring to the gas or vapor will carry the subscript v , whereas the unsubscripted symbols will refer to the particles. Also, while the coordinates and velocities will have the same nomenclature for both the solid and liquid particles, their normalizations will differ; these are specified in the Nomenclature.

It is shown in Ref. 1 that in the flow of particulates in a gaseous stream the pressure forces are small as compared to the inertia and viscous forces and, therefore, the general equation for the flow of a particle in a gaseous stream is given by:

$$m (dU/dt) = F = C_1(U_v - U).$$

From Ref. 2, $C_1 = 3\pi \mu_v D$, and since $m = 1/6 \rho \pi D^3$, we have:

$$C_p (dU/dt) = (U_v - U).$$

The three momentum equations in a cylindrical coordinate system thus become:

$$\frac{du}{dt} = \left(\frac{u_v - u}{C_p} \right) - \frac{uv}{r} \quad (1a)$$

$$\frac{dv}{dt} = \left(\frac{v_v - v}{C_p} \right) + \frac{u^2}{r} \quad (1b)$$

$$\frac{dw}{dt} = \left(\frac{w_v - w}{C_p} \right) \quad (1c)$$

Referring to Figure 1* for the general shape of the container geometry, we normalize the above expressions by the radius R_c and inlet velocity U_I to obtain:

*For ease of readership, figures and tables are presented consecutively at the end of the paper.

$$\frac{d\bar{u}}{d\tau} = \left(\frac{\bar{u}_v - \bar{u}}{\bar{C}_p} \right) - \frac{\bar{u}\bar{v}}{\bar{r}} \quad (2a)$$

$$\frac{d\bar{v}}{d\tau} = \left(\frac{\bar{v}_v - \bar{v}}{\bar{C}_p} \right) + \frac{R_c \omega}{U_I} \frac{\bar{u}^2}{\bar{r}} \quad (2b)$$

$$\frac{d\bar{w}}{d\tau} = \left(\frac{\bar{w}_v - \bar{w}}{\bar{C}_p} \right) \quad (2c)$$

The velocities, of course, are related to the particle position coordinates, $P(\theta, r, z)$, via $u = r(d\theta/dr)$; $v = (dr/dt)$; and $w = (dz/dt)$.

In Equation 2, the quantities u_v , v_v , w_v are input functions for the gas velocity field in the cycloentrifuge as obtained from a separate analysis, or any arbitrary fluid field one chooses to introduce. Equation 2 is then solved for particle position, $P(\theta, r, z)$, as a function of time from the $t = 0$ instant of its injection.

PARTICLE RICOCHET

When the solid particles are injected at the top of the cycloentrifuge the heavier pieces will go out via the bottom end at $z = L$, others will be carried out with the gas stream via the rotor at $z = L_1$. Some particles, before being discharged at either passage, will collide with the walls. As a consequence, they will lose some of their momentum and have their velocity altered in both magnitude and direction. The radial velocity of such an impinging particle will be altered by what is called the coefficient of restitution C_R , while the other two velocity components will be reduced due to the drag at the walls via a velocity coefficient C_V . The coefficient of restitution depends on the elastic properties of particles and wall, whereas C_V depends on the coefficient of friction. Accordingly, the rebound velocity is given by:

$$\vec{U}_F = C_V (\vec{u} + \vec{w}) - C_R \vec{v} \quad (3)$$

In the region $L_2 < z < L$ the radial velocity, v , is not normal to the wall. To have Equation 3 apply, we have to formulate new radial and axial velocities v' and w' such that they are normal and parallel to the inclined walls of the centrifuge at $z > L_2$. Thus, we have:

$$\begin{aligned} v' &= -w \sin \alpha + v \cos \alpha \\ w' &= w \cos \alpha + v \sin \alpha \end{aligned}$$

The component u remains unchanged. Therefore, the velocities following ricochet are given by:

$$u_F = C_V u \quad 0 < z < L \quad (4a)$$

$$v_F = -C_R v' \quad 0 < z < L_2 \quad (4b)$$

$$w_F = C_V w' \quad (4c)$$

or

$$v_F = v[C_V \sin^2 \alpha - C_R \cos^2 \alpha] + w(C_V + C_R) \sin \alpha \cos \alpha \quad (4d)$$

$$w_F = w[C_V \cos^2 \alpha - C_R \sin^2 \alpha] + v(C_V + C_R) \sin \alpha \cos \alpha \quad L_2 < z < L \quad (4e)$$

The set of differential equations [2] was solved on a computer essentially by a Runge Kutta method. The scheme was refined by the introduction of a "predictor corrector" algorithm which reduced considerably the time required to obtain solutions. The method is described elsewhere [3] and will not be gone into here.

SAMPLE SOLUTIONS

To give an idea of the path a solid particle is likely to take when injected into a particular flow field, trajectories were plotted for solid particles caught in the flow field of a cycloentrifuge. This device differs from an ordinary cyclone in that it has an inner rotor that imparts momentum to the injected gas. The dimensions and operating conditions of the prototype cycloentrifuge are given on Table 1. Figure 2 shows a typical flow field in the (r,z) plane of the gas, with most of it flowing out the top and a small portion via the bottom of the device. The solid particles, due to their inertia, will follow stream lines different from those of the gas, and it is these trajectories that are shown in the plots that follow.

In Figure 3, showing the trajectory for the smallest size, the particle essentially followed the path of the gas and as soon as it cleared the rotor exited via the center. The next size particle, that given in Figure 4, was caught between the two possible exits and remained in a stagnant orbit at some distance from the rotor exit. The 2- μ m particle (Figure 5) exited the bottom after experiencing some ricocheting at the beginning and end of its paths, whereas the largest particle, given in Figure 6, bounced against the outer walls of the cycloentrifuge all the way from injection to exit. In essence, the 3- μ m particle rubbed against the walls throughout its residence in the chamber with undesirable consequences for the integrity of the device. The most desirable size in the present conditions is the 2- μ m particle, which avoided excessive collisions with the walls and yet was separated from the gaseous stream, as intended.

DROPLET-GAS FLOW

EQUATIONS FOR LIQUID-GAS PARTICLES

Since the enclosure within which the particles flow, namely the blade spacing is itself rotating, the equations will be written in a rotating coordinate system, as shown in Figure 7. All velocities are relative to the rotating blades and the other bounding surfaces. However, as is clear from Figure 7, the radial and axial velocities are also absolute velocities, whereas the relative circumferential velocity, u , differs from the absolute circumferential velocity by the quantity $r\omega$.

In this scheme the three relevant momentum equations are:

$$\rho V_v \frac{d^2 \theta}{dt^2} = F_\theta - 2\rho V \left(\omega + \frac{d\theta}{dt} \right) \frac{dr}{dt} \quad (5a)$$

$$\rho V \frac{d^2 r}{dt^2} = F_r + \rho V_r \left(\omega + \frac{d\theta}{dt} \right)^2 \quad (5b)$$

$$V \frac{d^2 z}{dt^2} = F_z \quad (5c)$$

Above, F is the viscous drag force, the last term in Equation 5a is the Coriolis force, and the last term in Equation 5b is the centrifugal force, both imparted on the particle by the drag of the rotating fluid. The viscous force, F , can in general be represented by:

$$F = C_D \times (1/2) \rho A U_R^2$$

where the coefficient C_D is a function of the Reynolds number. Writing:

$$\bar{F} = \frac{F}{(1/2) \rho r_2 \omega^2} = 3C_D \left(\frac{\rho v}{\rho} \right) \left(\frac{r_2}{R_0} \right) \bar{U}_R^2$$

$$r = l\omega$$

we obtain the normalized form of Equation 5:

$$\bar{r} \left(\frac{d^2 \theta}{d\tau^2} \right) = \frac{\bar{U}_R C_D}{\bar{R}} (\bar{u}_v - \bar{u}) - 2 \left(1 + \frac{d\theta}{d\tau} \right) \bar{v} \quad (6a)$$

$$\frac{d^2 \bar{r}}{d\tau^2} = \frac{\beta \bar{U}_R C_D}{\bar{R}} (\bar{v}_v - \bar{v}) + \bar{r} \left(1 + \frac{d\theta}{d\tau} \right)^2 \quad (6b)$$

$$\frac{d^2 \bar{z}}{d\tau^2} = \frac{\beta \bar{U}_R C_D}{\bar{R}} (\bar{w}_v - \bar{w}) \quad (6c)$$

where:

$$\bar{U}_R = [(\bar{u}_v - \bar{u})^2 + (\bar{v}_v - \bar{v})^2 + (\bar{w}_v - \bar{w})^2]^{1/2} \quad (7)$$

$$\beta = \frac{3}{2} \left(\frac{\rho v}{\rho} \right) \left(\frac{r_2}{R_0} \right)$$

EVAPORATION OF LIQUID PARTICLE

In all the previous equations the size of the droplet, R , is a variable and a function of time. The variation in droplet size can be obtained from a heat balance between the heat required for evaporation and the rate of heat transfer occurring between the gas and droplet. The amount of heat required to induce a given rate of vaporization is:

$$\frac{dq}{dt} = h_{fg} \frac{dm}{dt} = (g\rho)h_{fg} \frac{dV}{dt} \quad (8)$$

Above, it is assumed that the liquid injected into the gas is at a temperature corresponding to the saturation temperature of the local pressure, and the only heat required is the vaporization enthalpy, h_{fg} .

The heat required for vaporization has to be transferred from the gas to the liquid particle by conduction, or:

$$dq/dt = \lambda A (T_v - T_s) \quad (9)$$

Writing $A = 4\pi R^2$, $V = (4/3) \pi R^3$, and equating the two rates of heat transfer, we obtain:

$$\frac{dR}{dt} = \frac{\lambda(T_v - T_s)}{g\rho h_{fg}} \quad (10)$$

Now the film coefficient, λ , can be related to the heat conductivity between the vapor and water droplet [4], via, $\lambda D/k = 0.37 Re^{0.6}$.

The above inserted into Equation 10 and properly normalized, yields:

$$\bar{R} \frac{d\bar{R}}{d\tau} = -\gamma Re^{0.6} [T_v - T_s] \quad (11)$$

where:

$$\gamma = \frac{0.74 k}{g\rho D_o^2 \omega h_{fg}}$$

If the variation in T_v is not very large then γ is approximately constant, with only Re and ΔT varying. If T_v varies over a wide range then certainly h_{fg} , and to some degree k , would vary and Equation 11 should more properly be written as:

$$\bar{R} \frac{d\bar{R}}{d\tau} = -\frac{0.74}{g\rho D_o^2 \omega} \left[\frac{k Re^{0.6} (T_v - T_s)}{h_{fg}} \right] = -\frac{0.74}{g\rho D_o^2 \omega} F(\theta, r, z) \quad (12)$$

The above differential equation for $R(\tau)$ must be added to of Equation 6 for a complete solution of the problem of droplets evaporating in a high-speed fluid. These equations constitute an initial value problem and need starting conditions in order to be solved. These are supplied by the mode of droplet injection at $t = 0$, namely:

Point of injection θ_0, r_0, z_0
 Velocity of injection u_0, v_0, w_0
 Particle size R_0
 Relevant constants Re_0, β, γ

PARAMETRIC SOLUTIONS

Two sets of solutions will be presented here. The first serves as a brief introductory survey of the basic behavior of droplet flow in a rotating system. The second and main series deals with an actual compressor whose characteristics are described in detail in Ref. 1.

Gas with Radial Flow Only

In order to bring out some of the basic features of droplet behavior, a preliminary run was made under the following conditions:

$$u_v = u_v = w_v = 0$$

$$\bar{v}_v = 0.5 ; u_v = w_v = 0$$

The first represents a case where the gas moves integrally with the blades. The results show that the drag imposed on the droplet tends to keep it at a nearly constant radius, in phase with the gas. Thus, the Coriolis and centrifugal forces seem to play a small part as compared to the viscous drag, at least for the $10 \mu\text{m}$ particle considered in this example. The second set of conditions represents a situation of constant gas velocity in the radial direction only. Small particles are completely dominated by the viscous drag; however, $30 \mu\text{m}$ particles have sufficient inertia to remain in the flow field until they are caught up and struck by the upstream blade.

The rate of change in particle diameter due to evaporation, as shown in Figure 8, seems to go through three stages. It is high at the beginning of injection, probably because at that point the relative velocity is high; it then tapers off, to increase again toward the end of the path. The latter increase is due to the decrease of particle size, which, as will be seen later on, yields higher rates of evaporation. Since the variation in mass is proportional to $(1 - R)^3$, the above variation in slope will be even more pronounced for the particle's rate of change of mass.

Three-Dimensional Gas Flow

In schematic form the blade geometry in which the flow takes place is given in Figure 9. Three particle sizes were investigated: 2, 10, and $50 \mu\text{m}$.

Figures 10 through 14 show the dynamics of the droplet in a three-dimensional gas flow field. While the velocity map of the gas is specific as

detailed in Ref. 1, the parametric behavior of the droplet is general and should hold for any other underlying flow map be it turbine or compressor. The results shown in the figures are summarized in the following paragraphs.

Injection from Pressure Side -- Physically this mode of injection means that the particle has little opportunity to move away from the plane of injection. As a result, small particles whose inertia is low, in most cases are struck by the pressure side of the blade; therefore, no results are offered for particle sizes of 2-10 μm . As shown in Ref. 10 for $D_0 = 50 \mu\text{m}$, increasing the u_0 velocity from 0.2 to 0.5 did not seem to make much difference. Likewise, runs in which the angle of injection was varied from 90 to 45° (i.e., with $v_0 = \pm u_0$), did not improve the exiting chances, and in some cases made the situation worse.

Injection from Suction Side -- Injection against rotation (i.e., with the injection plane moving away from the particle), provides satisfactory trajectories, enabling the particles to exit from the chamber. Here, as seen from Figure 11, lighter particles are preferred, as with the $D_0 = 50 \mu\text{m}$ particles the vapor does not succeed in sweeping them out the channel before they are struck by the approaching upstream blade. The very light particles are swept out of the channel along a path nearly parallel and very close to the injection plane. Here, in contrast to the situation with $\theta_0 = 0$, the lower the injection velocity the better the exiting chances. Thus, as seen in Figure 12, by reducing the injection velocity to nearly zero, and placing the injection point at $\bar{z}_0 = 0.55$ even the $D_0 = 50 \mu\text{m}$ particles succeed in exiting. For the large particles collision with the pressure side of the blade occur whenever $|u_0| > 0.55$. The effects of varying particle size are clearly shown in Figure 13, in that collisions with the walls can be avoided with small size particles.

Effect of Injection Angle -- We shall consider the effects of varying the injection angle, while keeping the total velocity constant. For instance:

$$\bar{U}_0 = [\bar{u}_0^2 + \bar{u}_0^2 + w_0^2]^{1/2} = 0.2$$

In discussing the results for various injection angles we shall maintain the following convention:

1. $\phi_\theta = 0$ is the injection normal to the suction plane or $\bar{U}_0 = -0.2$
2. $\phi_\theta = 45^\circ$ refers to a velocity vector tilted in the $r\theta$ plane 45° toward the exit (i.e., toward positive r)
3. $\phi_\theta = -45^\circ$ is the same as above but with a tilt away from the exit ($-r$)
4. $\phi = 45^\circ$ refers to a velocity vector tilted in the (θz) plane 45° toward the hub ($+z$)
5. $\phi = -45^\circ$ is the same as above with a tilt toward the shroud ($-z$).

For small particles $D_0 \approx 2 \mu\text{m}$, varying the injection angle has no effect whatsoever. Also, the vaporization rates are little affected by varying the orientation of the velocity vector. Furthermore, as shown in Figure 14, the

effect of varying ϕ_0 on the droplet trajectories is also minor. The effects of orienting the flow 45° either toward the hub or toward the shroud are shown in Figure 15. First, it will be noted that for either plus or minus values of ϕ_z , the trajectories of the particles below $10 \mu\text{m}$ are very close to that of normal injection. However, there are important effects on the trajectories of the heavier particles. For $\phi = +45^\circ$ the result is that the larger particles collide with the wall of the hub. However directing the flow toward the shroud, $\phi = -45^\circ$, produces significant excursion away from the $\phi = 0$ solution.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and, therefore, the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

NOMENCLATURE

CC	=	Cyclocentrifuge	r_2	=	Blade outer radius
C_D	=	Viscous drag coefficient	$\bar{r}$	=	(r/r_2) or (r/R_c)
C_R	=	Coefficient restitution	t	=	Time
C_v	=	Velocity coefficient	t_0	=	R_c/U_I
C_p	=	$D^2/18 \nu_v$	u	=	Circumferential velocity
$\bar{C}_p$	=	$D^2 U_I / 18 R_c \nu_v$	$\bar{u}$	=	$(u/r_2 \omega)$ or $(u/R_c \omega)$
C_p/t_0	=	C_p	v	=	Radial velocity
D	=	Diameter of particle	$\bar{v}$	=	$(v/r_2 \omega)$ or (v/U_I)
F	=	Force	w	=	Axial velocity
L	=	Height of centrifuge	$\bar{w}$	=	$(w/r_2 \omega)$ or (w/U_I)
$L_{1,2}$	=	See Figure 1	z	=	Axial coordinate
N	=	Rotational frequency	$\bar{z}$	=	(z/r_2) or (z/R_c)
Q	=	Mass flow	β	=	$(3/2)(\rho_v/\rho)(r_2/R_0)$
R	=	Radius of particle	γ	=	$0.74k/\rho g D^2 \omega \eta g$
$\bar{R}$	=	(R/R_0)	θ	=	Angular coordinate
R_c	=	Radius of cyclocentrifuge	λ	=	Film heat transfer coefficient
$R_{1,2}$	=	See Figure 1	μ	=	Absolute viscosity
Re	=	Reynolds number, $Re_0 \bar{U}_R \bar{R}$	μm	=	Microns, 10^{-6}m
Re_0	=	Reference Reynolds number, $D_0 r_2 \omega / \nu_v$	ν	=	Kinetic viscosity
T	=	Absolute temperature	ρ	=	Density
U	=	Particle velocity relative to walls	τ	=	$t\omega$ or $t(U_I/R_c)$
			$\Delta\tau$	=	Residence time of droplet until exit or impact

U_I	= Inlet velocity to centrifuge	ϕ	= Angle between z-axis and streamline
U_R	= Particle velocity relative to gas	ω	= Angular velocity
$\bar{U}$	= $U/r_2\omega$		
V	= Volume	<u>Subscripts</u>	
c	= Specific heat	c	= Cyclocentrifuge
g	= Gravitational constant	F	= Final degrees
k	= Heat conductivity	o	= Point of injection
m	= Mass	s	= Saturation
q	= Heat	v	= Vapor or gas
r	= Radial coordinate		

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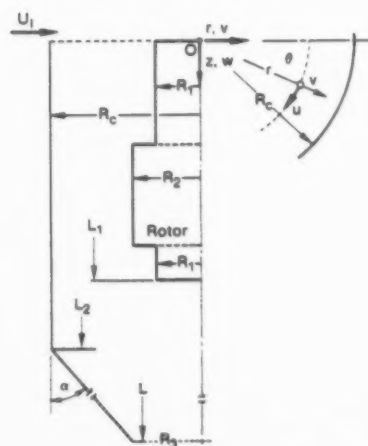


Figure 1. Schematic of cyclocentrifuge.

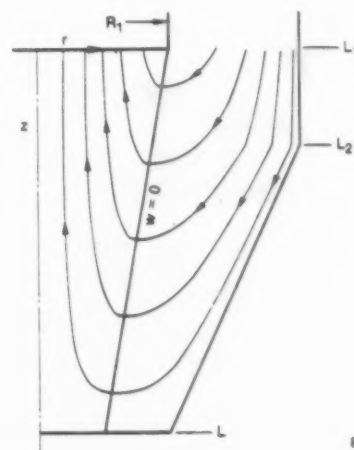
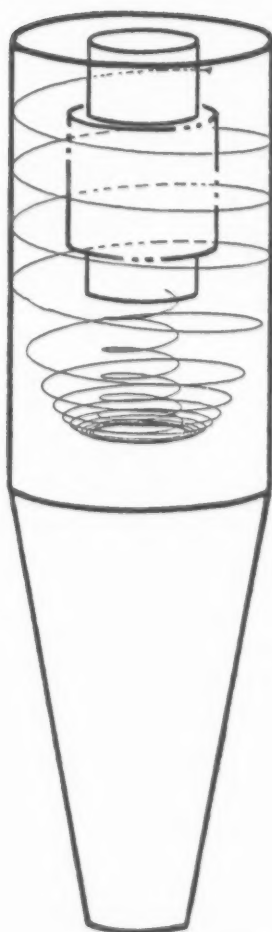


Figure 2. Gas streamlines in cyclocentrifuge.



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Figure 3. Particle trajectory
for $d = 0.5 \mu\text{m}$.



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Figure 4. Particle trajectory
for $d = 1.5 \mu\text{m}$.

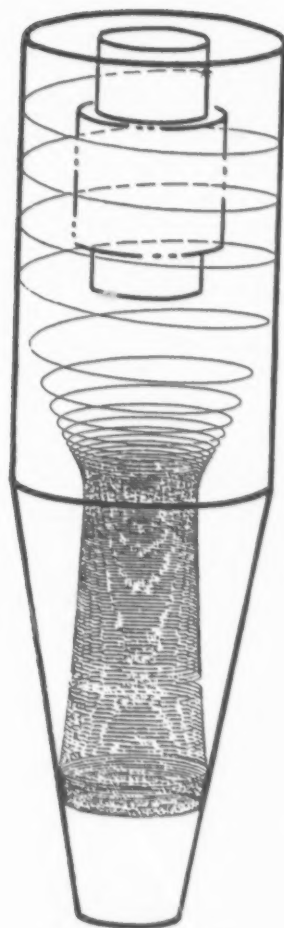


Figure 5. Particle trajectory
for $d = 2 \mu\text{m}$.

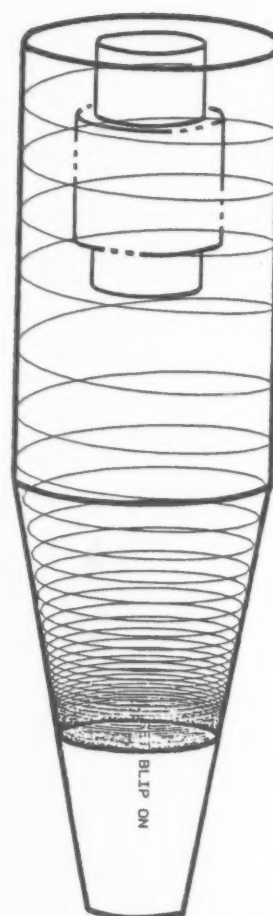
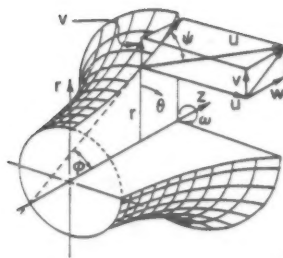
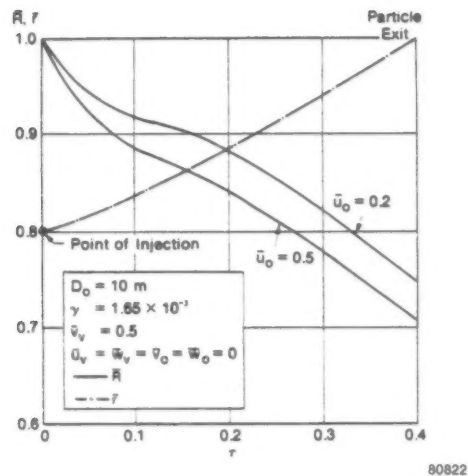


Figure 6. Particle trajectory
for $d = 3 \mu\text{m}$.



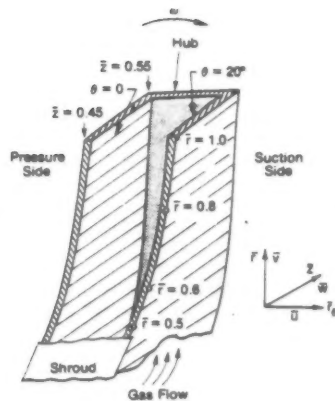
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Figure 7. Coordinate systems for droplet gas flow.



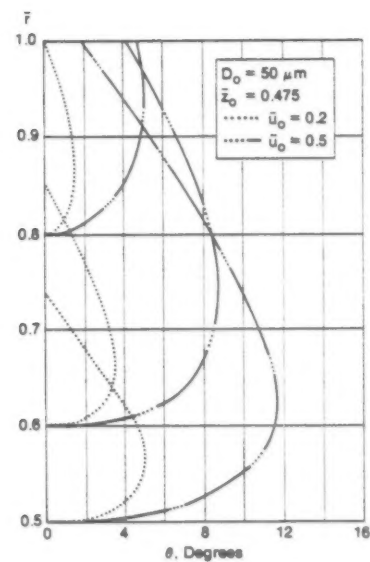
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Figure 8. Droplet flow in a gas with radial velocity only.



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Figure 9. Blade space geometry and nomenclature.



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Figure 10. Injection from high-pressure side.

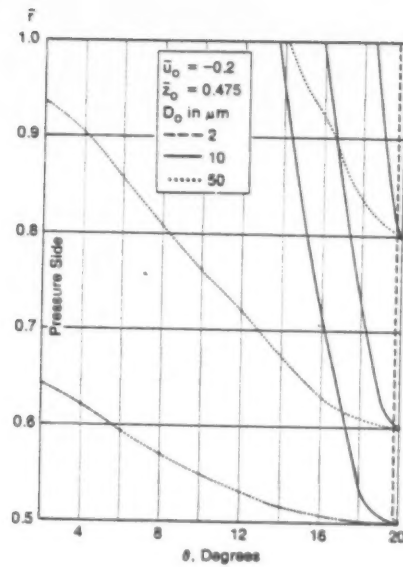


Figure 11. Injection from low-pressure side.

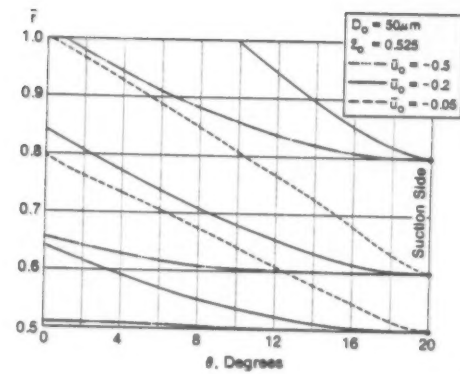


Figure 12. Effect of injection velocity.

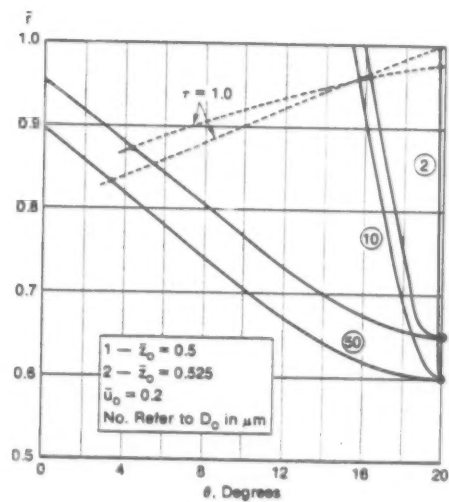


Figure 13. Effect of particle size.

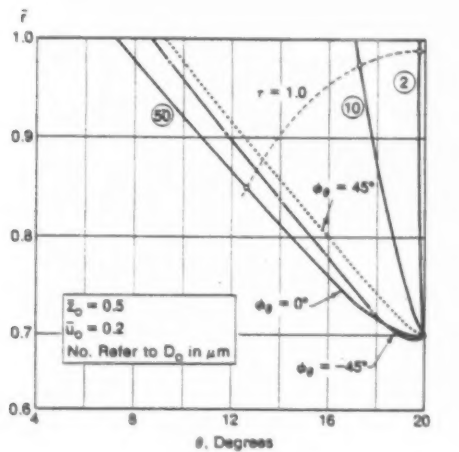


Figure 14. Effect of injection angle in $r\theta$ plane.

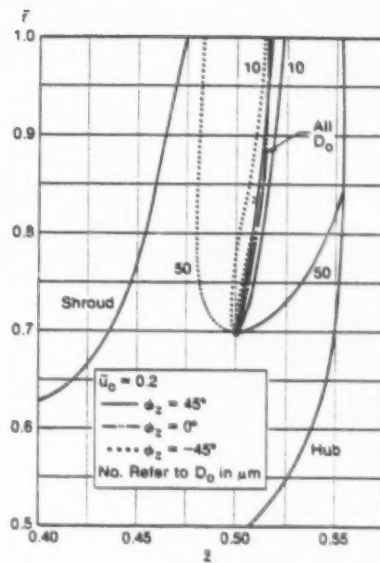


Figure 15. Effect of injection angle in rz plane.

Session C-5: INTEGRATED CONTROL PROCESSES I

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PARTICULATE CONTROL INTEGRATION WITH
WET SCRUBBER SO₂ REMOVAL AND
WASTE GENERATION

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Brown and Caldwell Engineers,
Steven M. Katzberger, Sargent and Lundy Engineers,
J. Edward Cichanowicz, EPRI

ABSTRACT

This paper presents the results of tests of fabric filter/wet scrubber and electrostatic precipitator/wet scrubber environmental control configurations from the EPRI 2.5 MW (e) Integrated Environmental Control Pilot Plant (IECPP). Higher particulate penetration from the ESP limited SO₂ removal by restricting the pH control point; this occurred when the limestone scrubber was operated under closed loop water management conditions. The presence of aluminum and fluoride was identified as the cause of these operating problems. The aluminum is probably leached from the ash; fluoride is scrubbed from the flue gas. These control and removal problems could be avoided by (1) operating under open loop water management, (2) using lime rather than limestone or (3) reducing particulate penetration to the scrubber.

INTRODUCTION

The Integrated Environmental Control (IEC) program in Electric Power Research Institute's (EPRI's) Coal Combustion Systems Division was established to develop a systems design approach for flue gas emissions control, and techniques for management of wastewater, waste solids, and waste heat. The IEC program was established to address the increasing complexity of coal-fired power plant environmental controls due to continually evolving regulations for flue gas emissions and wastewater, waste solids, and waste heat. This increase in environmental control complexity has been a major factor in the escalation of capital costs and operating penalties of availability and heat rate.

The IEC program consists of engineering studies and pilot plant tests to: (a) identify concepts for simplifying control system design, and (b) avoid additional operating penalties as control requirements for wastewater and waste solids are increased. Pilot plant tests, which are conducted using the Integrated Environmental Control Pilot Plant (IECPP), have two objectives:

- Determine the interactions between emission control components. These include gas controls, both particulate and gaseous; water management, primarily closed versus open loop; and solid waste disposal, including combined or separate disposal of fly ash and scrubber sludge.
- Learn how to use interactions between control components. This objective includes exploiting beneficial effects to simplify control system design and avoiding detrimental effects which could increase operating costs. Beneficial interactions observed have been an increase in thickener waste settling rates when reusing cooling tower blowdown (CTB) for limestone scrubber makeup. The reuse of CTB for scrubber makeup eliminates or reduces wastewater disposal requirements. Detrimental interactions have been observed when particulate penetration levels have increased aluminum and fluoride levels affecting limestone dissolution, pH operating point, and ultimately SO₂ removal.

Pilot-scale testing of IEC configurations compares the efficiency and performance of the components of each configuration. The IECPP is located at the EPRI Arapahoe Test Facility in Denver, Colorado. The test facility is immediately adjacent to, and receives flue gas from, Colorado Public Service Company's Arapahoe Power Station Unit 4.

Presently, four generic environmental control configurations will be tested. The configurations use the two most common particulate control devices, the electrostatic precipitator (ESP) and fabric filter (FF), with the two most common SO₂ control systems, wet scrubbing (WS) and spray drying (SD). The IECPP has the flexibility to use either lime or limestone for reagent and the facilities to prepare makeup water to simulate cooling tower blowdown or other plant wastewater streams to test various water management reuse options. Solid waste samples from the particulate control and SO₂ control devices are mixed and tested for strength and permeability.

The status of the pilot plant test schedule is presented in Figure 1. The FF/WS tests were completed in December 1983. The ESP/WS tests were completed in December 1985. The SD/ESP and SD/FF tests are planned for 1986. Dry additive injection tests using the IECPP are being considered for 1987. This paper presents results from the first two configurations. The test program, pilot plant, and results are discussed in the following sections.

TEST PROGRAM

The specific tests for each configuration were defined after considering the design and operating conditions for each configuration. The logic behind the development of the specific test conditions is presented in this section.

Figure 2 shows the distribution of FF/WS and ESP/WS configurations throughout the United States. There are 118 operating ESP/WS and 6 operating FF/WS with 3 FF/WS systems in various planning or construction stages. All of the FF/WS are (or will be) located in the West, whereas the ESP/WS are located throughout the United States. The geographical restriction of the FF/WS can be explained by two events. First, fabric filters have only recently become acceptable control devices for utility boilers, so only newer plants or recent upgrades have them. Second, the comparative economics of fabric filters versus ESPs favor fabric filters when a unit is firing low sulfur, high moisture western fuel. The high resistivity of the fly ash from this fuel requires a very large specific collecting area (SCA) (600 to 1,000 ft²/acfm) and thus large, and expensive precipitators. With midwestern and eastern fuels, the higher sulfur content lowers fly ash resistivity and smaller, less expensive precipitators are often more economical than fabric filters.

Figure 3 shows the net evaporation and precipitation areas in the United States along with the distribution of the FF/WS and ESP/WS. This figure shows that the FF/WS systems are located in net evaporation areas where the power plants are generally operated under closed loop water management. Another similarity among the FF/WS systems is that they have design SO₂ removal requirements of approximately 80 percent because of their use of low

STATUS	CONFIGURATION
COMPLETE	NO. 1 - FABRIC FILTER/WET SCRUBBER
COMPLETE	NO. 2 - ELECTROSTATIC PRECIPITATOR/ WET SCRUBBER
APRIL - NOVEMBER 1986	NO. 3 - SPRAY DRYER/ESP
JANUARY - MARCH 1987	NO. 4 - SPRAY DRYER/FABRIC FILTER
MAY - JULY 1987	DRY INJECTION - SODIUM BASED, HYDRATED LIME - ESP AND FF

Figure 1 Schedule of IECPP Test Programs



- PRECIPITATOR AND WET SCRUBBER
- ▲ FABRIC FILTER AND WET SCRUBBER
- FABRIC FILTER AND WET SCRUBBER (PLANNED)

Figure 2 Distribution of Fabric Filter Wet Scrubbers and ESP Wet Scrubbers in the United States

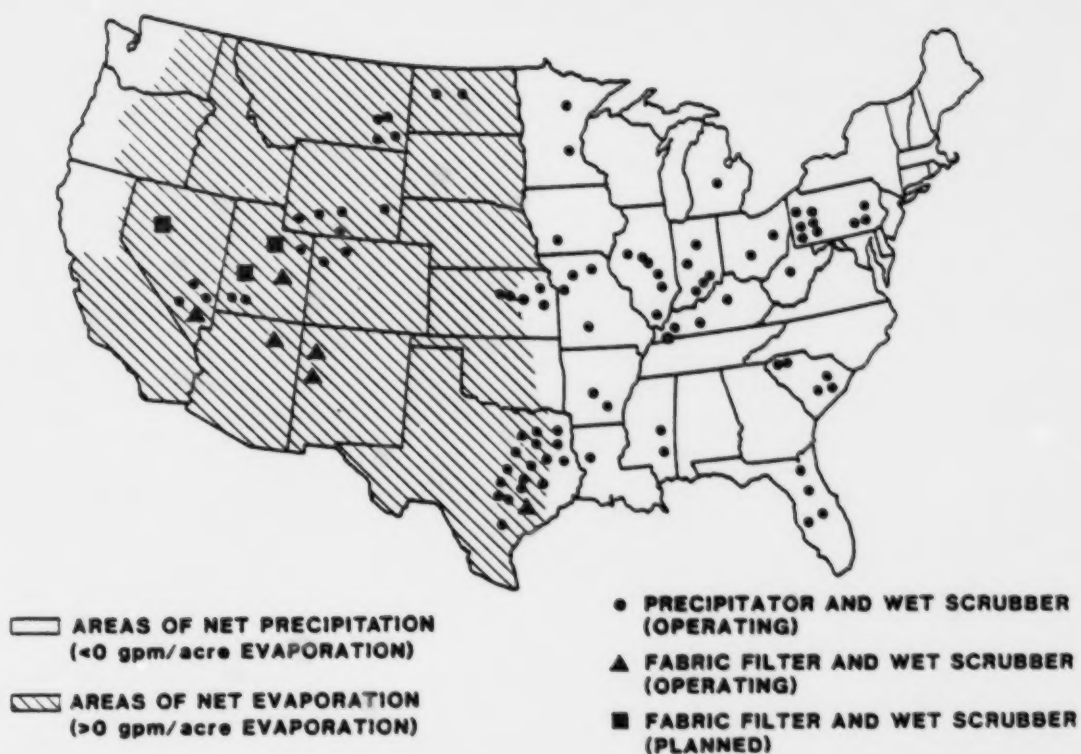


Figure 3 Annual Precipitation and Evaporation in the United States

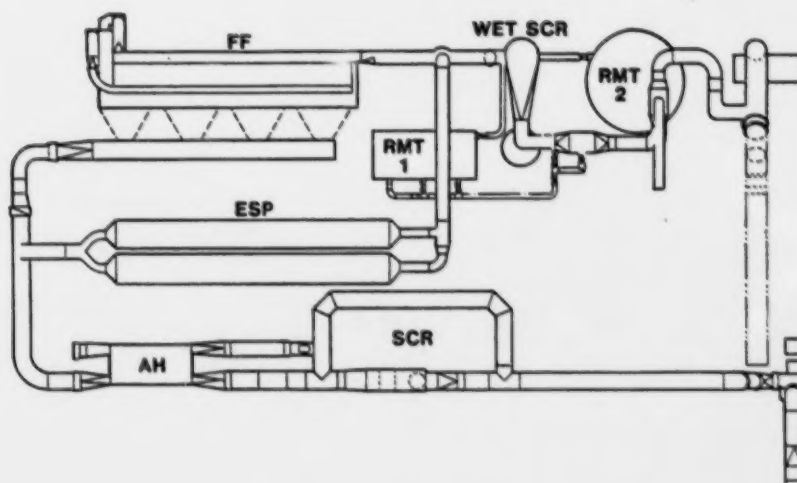


Figure 4 Schematic of IECPP Configurations 1 and 2

sulfur coal. The ESP/WS systems are found on units firing fuels with various sulfur levels and SO₂ collection efficiencies ranging from 70 to 95 percent. Consideration of the distribution of ESP/WS and FF/WS systems, and water management and SO₂ removal requirements led to the development of the test programs. The selected test conditions are briefly discussed below.

The major design criteria for FF/WS systems include: zero (or severely limited) discharge of plant water streams, relatively low (70 to 80 percent) SO₂ removal; and modest, but increasingly restrictive, requirements on the disposal of waste solids. The test plan for the FF/WS assumed that this system would be located primarily in the western United States, and because very few FF/WS configurations are in operation, the test plan was directed to new plant design concepts. In contrast to this, the majority of ESP/WS systems are located east of the Rocky Mountains where coals with lower resistivity ash are burned. Since ESPs have historically removed particulates and because retrofit SO₂ control requirements are possible in the future, both retrofit and new plant design considerations were considered. The problems inherent in locating waste disposal sites for retrofit scrubber systems make waste disposal a major concern for ESP/WS retrofit designs. These facilities may have very restrictive requirements on solid waste disposal. Additionally the location of the ESP/WS systems allows the option of open loop water management. The primary differences between test conditions of Configurations 1 and 2 are:

- The plants which have an ESP/WS control system are not required to meet zero discharge design criteria because they are located in areas of net precipitation. These plants are permitted to discharge cooling tower blowdown, decreasing emphasis on using CTB in the scrubber.
- Land for disposal of scrubber sludge from ESP/WS retrofit application may not always be readily available, requiring low sludge moisture content to reduce haul costs.
- Solid waste management criteria for ESP/WS retrofit systems may be more strict, requiring in some instances, adequate strengths (>50 psi) and low permeabilities (<10⁻⁶).
- There will be more particulate penetration to the wet scrubber from ESP/WS systems which will affect solids management and disposal considerations. Particulate control levels for the ESP/WS SO₂ were varied to simulate application of either new units meeting 1979 NSPS, (0.03 lb/MMBtu) or retrofit meeting 1971 NSPS (0.1 lb/MMBtu) or earlier standards with emissions in the 0.2 lb/MMBtu range.

The test program for the FF/WS emphasized tight water management, use of freshwater and simulated cooling tower blowdowns for scrubber makeup, different air heater exit temperatures (simulating the influence of waste

heat recovery systems), at high (2,000 ppm) and low (300 ppm) SO₂ levels. The high SO₂ tests were included to determine the sensitivity of the results to SO₂ concentration and to minimize the time to reach steady-state in the scrubber system.

The proposed test program for the ESP/WS deemphasized the use of wastewater as makeup for the scrubber, evaluated different particulate emission levels from the ESP, considered forced oxidation in the scrubber, and emphasized high SO₂ tests. As test results became available and the importance of particulate carryover was established, the test plan was changed to include closed- and open-loop water management and different reagents. The reasons for the change in this test program are discussed after a brief discussion of the pilot plant which follows.

INTEGRATED ENVIRONMENTAL CONTROL PILOT PLANT

The pilot plant receives flue gas from Unit 4 of the Arapahoe Power Station operated by Public Service Company of Colorado. The boiler is a B&W, vertically-fired unit with a capacity of 100 MW. The unit burns Colorado and Wyoming coals which have low sulfur and medium heating values. The plant is located in Denver, Colorado, at an elevation of approximately 5,300 feet. A schematic of the IECPP is shown in Figure 4. A brief description of the major components of interest is presented below.

FABRIC FILTER

The IEC fabric filter is a four-compartment Ecolaire unit. Table 1 gives design and operating characteristics. The air-to-cloth ratio is 2.61:1, and the performance of the fabric filter reflects this conservative design. The fabric filter has several different cleaning mechanisms but reverse gas cleaning was used, sometimes supplemented by shaking or horns. Several characteristics of this baghouse differ significantly from full scale units. First, this unit exhibits high temperature losses of 30 to 40 degrees F due primarily to the high surface-to-volume ratio. Second, the residence time is almost 40 seconds, because the volume of the unit includes a 2-foot wide walkway around the perimeter of the tubesheet resulting in a large volume per compartment. Third, in-leakage of air along door seams is 5 to 10 percent. All of these factors contributed to low outlet temperatures (160 degrees F).

ELECTROSTATIC PRECIPITATORS

The IECPP uses two four field electrostatic precipitators in parallel. The design and operating characteristics are presented in Table 2. The SCA at 11,000 acfm is only 366 ft²/1,000 acfm. Based on tests with similar pilot plant units at the Arapahoe test facility, particulate emission levels were expected to be between 0.03 and 0.15 lb/MMBtu, depending on fly ash

Table 1. FABRIC FILTER DESIGN AND OPERATING CHARACTERISTICS

<u>Description</u>	<u>Value</u>
Dimensions (approximate)	
Length	42 feet
Width	21 feet
Height	51 feet
Number of compartments	4
Number of bags per compartment	36
Total number of bags	144
Gross filter area	6,668 sq ft
Net filter area (1 compartment cleaning)	5,001 sq ft
Service filter area (1 compartment cleaning and isolated)	5,001 sq ft
Air-to-cloth-ratio	
Gross	1.61:1
Net	2.61:1
Design volume	10,800 acfm
Operating temperature	
min/design/max	170/300/500
<u>Screw Conveyor System</u>	
One (1) gather up conveyor	
Diameter	9 inches
Length	38 feet
Type	U-trough/top loading
<u>Reverse Gas Fan</u>	
Type	
Volume	Zurn-Clarage "XL"
Static pressure @ 500 degrees F	8,500 acfm
	12 inches W.C.
<u>Shaker System</u>	
Type	Tabular
Motor hp	1
Motor rpm	1,800
Shake frequency	300 cpm
<u>Filter Bags</u>	
Fiber	Glass
Weave	3 x 1 twill
Count	54 x 30
Weight (oz/yd)	9.6
Permeability	60-75
Finish	Silicon - graphite Teflon
Diameter	8 inches
Length	23.8 ft
<u>Horns</u>	
Type	Fuller
Model	310-64-1-0099
Number/compartment	1
Location	Top
Orientation	Down
Air requirement	
Scfm	30
Psi	90

Table 2. ELECTROSTATIC PRECIPITATORS DESIGN AND OPERATING CHARACTERISTICS

Number of units	2
Arrangement	In parallel
Casing dimensions (each unit) (excluding hopper or insulator housings)	42 feet long x 4 feet wide x 8 feet high
Fields/ESP	4
Number of plates/field	5
Plate spacing distance	9 inches
Number of plate sections	6 (Each 18 inches long)
Overall plate dimensions	9 feet 6 inches long x 7 feet 6 inches high
Effective plate area	500 feet ² /field
Specific collection area	366 feet ² /1,000 ACFM at 11,000 ACFM
Plate rapping, field 1	2 revolutions of lifter shaft every 5 minutes
field 2	2 revolutions of lifter shaft every 10 minutes
field 3	2 revolutions of lifter shaft every 15 minutes
field 4	2 revolutions of lifter shaft every 15 minutes
Number of high voltage emitter harps	4/field
Wires/harp	10
Wire rapping	10 raps (one/second) every 15 minutes
Harp construction	0.1035 inch wire, tensioned in 1 1/4-inch pipe frame
High voltage emitter rapper (number)	2/field (front and back)
High voltage emitter rapper type	Direct current pulse "drop hammer"
Material handling	Single screw in each trough
	Dust emptied every 4 hours
Gas flow distribution baffles (each unit)	4 perforated plates inlet 1 perforated plate outlet
Automatic voltage controllers (AVC) (all eight fields in automatic spark control)	3 from MWL Inc. 5 from Joy Western
Transformer/rectifiers	6-NWL 5.52 KVA 2-NWL 33 KVA

Table 3. PLANT INPUTS

Limestone			Fuel			Ash		
Constituent	FF	ESP	Constituent	FF	ESP	Constituent	FF	ESP
CaCO ₃	94.87	94.87	Moisture	7.0	10.5	SiO ₂	52.8	62.8
MgCO ₃	0.75	0.75	Ash	7.1	7.8	AL ₂ O ₃	24.4	26.2
Insol.	3.23	3.23	Sulfur	0.3	0.4	TiO ₂	0.8	0.5
SiO ₂	3.15	3.15	Nitrogen	1.5	1.4	Fe ₂ O ₃	6.9	2.4
Fe ₂ O ₃	0.12	0.13	Carbon	68.9	63.5	CaO	5.2	4.6
AL ₂ O ₃	0.13	0.18	Hydrogen	4.8	4.4	MgO	1.0	1.3
Na ₂ O _x	0.18	0.007	Oxygen	10.2	12.0	Na ₂ O	0.9	1.1
H ₂ O	0.61	0.43	Chlorine	0.01	0.01	K ₂ O	1.0	1.1
F	0.02	0.022	Fluorine, ppm	50	54.2	P ₂ O ₅	1.3	0.7
Mn	0.62	0.015				SO ₃	3.3	0.3
Zn	0.01	0.013						

- Notes: 1. Values expressed in percent unless otherwise noted.
2. FF refers to FF/WS test conditions.
3. ESP refers to ESP/WS test conditions.

resistivity. The resistivity of the fly ash from Unit 4 can vary from 10^{10} to 10^{12} ohm-cm. An SO_3 conditioning unit is available to reduce the resistivity of the fly ash to lower particulate penetration levels when required.

WET SCRUBBER

The IEC wet scrubber system consists of a spray tower, two reaction mix tanks, slurry recirculating pumps, a horizontal and vertical (gas flow) mist eliminator, and a reheater. The scrubber tower is a counter current flow open tower with the capability of achieving L/G ratios from 0 to 120. The high L/G was intended to provide 90 percent removal at 2,000 ppm. The scrubber was designed to operate with both low and medium sulfur coals producing SO_2 gas concentrations of 300 or 2,000 ppm, respectively. The higher SO_2 gas concentrations are produced by the flue gas conditioning system. The lower SO_2 levels are unconditioned gas as received by the pilot plant from Unit 4.

TEST RESULTS

This section presents the results of the FF/WS and ESP/WS tests which highlight the interaction that particulate penetration can have on SO_2 removal. The results are presented for the flue gas emissions and water management tests. A summary of the solid waste tests is also included.

FLUE GAS EMISSIONS

The range of tests conducted in this phase of the program include high and low SO_2 tests for both the FF/WS and ESP/WS and tests with different particulate penetration levels for the ESP/WS.

Baseline Tests

Low SO_2 baseline tests with the FF/WS were conducted under the following test conditions: closed loop water management, L/G ratio of about 30 (gpm/1,000 acfm), a limestone stoichiometric ratio of 1.05, a pH control point of 5.4. Test results confirmed design requirements for low particulate penetration (.0006 lb/MMBtu) and low (70 percent) SO_2 removals at design conditions. The scrubber performance could be improved to more than 90 percent by changing the pH control point, or increasing L/G to 70.

The baseline test conditions for the ESP/WS were similar to those for the FF/WS except particulate carryover into the scrubber, 0.15 lb/MMBtu, was 200 times higher than during the FF/WS tests. The results of the ESP/WS tests were dramatically different and inconsistent with performance expected from field experiences with ESP/WS systems. At low SO_2 with closed loop water management, the limestone scrubber operating pH could not be raised

above 5. On some occasions, almost three times the stoichiometric limestone requirement was added to the system without any appreciable change in pH. Under these tests, the total solids level in the scrubber slurry was 14 percent with carbonate comprising 60 percent of these solids. These high carbonate levels affected the dewatering systems operation, plugging the thickener underflow and underflow storage tank but did not affect scrubber slurry pH.

After verifying that the pH meter and other scrubber monitoring devices were accurate, reagent quality was examined as the reason that pH could not be raised with limestone. After receiving a new supply of limestone and verifying its quality, and still not being able to raise the pH above 5, the reagent was eliminated as a source of the problem. We then examined the flue gas and makeup water to determine if changes in these plant inputs could be responsible for the operating problems. However, both the makeup water and flue gas quality were quite similar to conditions present during the FF/WS tests. A comparison of the limestone, fuel quality, and fly ash during the FF/WS and ESP/WS tests are presented in Table 3. Since plant inputs were essentially the same during the ESP/WS tests as the FF/WS tests, a series of tests were undertaken with different reagents to investigate their effects on pH control. The pH of the scrubber in the ESP/WS system was responsive to addition of lime, soda ash, and sodium bicarbonate. Hourly doses of lime would raise the pH above 6 but as the lime was consumed, the pH would drop to 5.0 despite the addition of limestone at a pH control point of 5.4 as shown in Figure 5.

Limestone blinding from sulfite interference was also considered as a potential source of the problem. However, this was eliminated after several tests when all reagent addition was stopped to consume the excess limestone. The system pH would eventually drop below 3 indicating that all the limestone was exhausted. Limestone addition would then resume and pH response would be excellent up to pH 5 where the system pH would stop increasing as long as only limestone was added. The system was restricted to operation at a pH of 5 with limestone as the reagent. This restriction affected peak SO₂ removals during the low SO₂ ESP/WS tests. A typical scrubber response curve at low SO₂ for the ESP/WS tests and the FF/WS tests are presented in Figure 6.

The results of the investigations led us to conclude that pH control problems were due to the use of an ESP and the attendant higher particulate loading compared with the FF/WS. The measured variables are summarized and compared with the FF/WS tests in Table 4. The significant differences are:

1. Oxidation. Almost complete oxidation occurred with the ESP as compared to nominally 50 percent with the fabric filter. This is due to the lower slurry pH and possibly to the presence of oxidation catalysts which could be generated from ash carried over from the ESP.

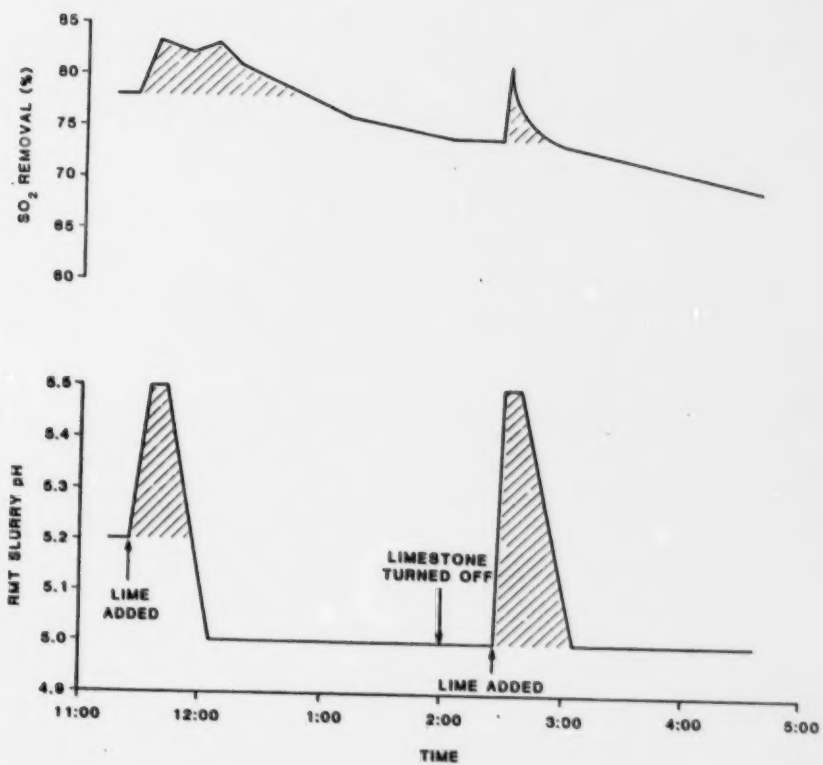


Figure 5 The Effect of Lime and Limestone Addition on pH Control and SO₂ Removal

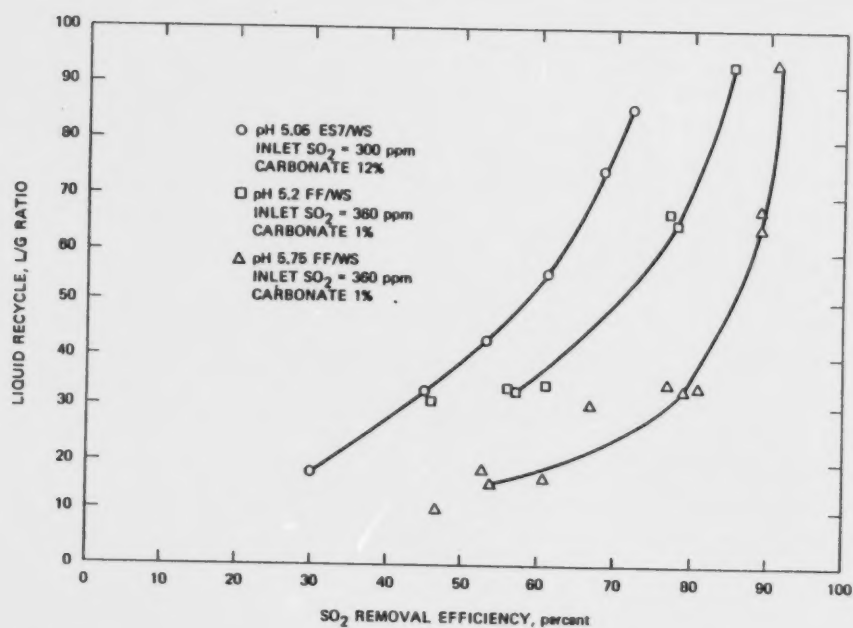


Figure 6 Scrubber Characteristic - Low SO₂ - ESP/WS Test

Table 4. COMPARISON OF BASELINE TEST RESULTS

Observation	Fabric filter/wet scrubber	ESP/wet scrubber
System chemistry		
Oxidation, percent	50	100
Aluminum, mg/l	Not detected	100-200
Fluoride, mg/l	10-15	500-700
Dissolved solids, mg/l	10,000	15,000
Calcium	1,380	713
Carbonate, percent	0-82	10-60
Particulate levels into scrubber		
Gr/dscf	0.0002	0.05
lb/MMBtu	.0005	0.1-0.15
SO ₂ removal, percent	75	60-74
Test conditions		
Inlet particulate		
Gr/dscf	3.1	2.7
L/G	30	30
pH	5.4	5.0-5.1
Inlet SO ₂	300	305
Reagent	Limestone	Limestone
Stoichiometric ratio	1.1	1.1
Thickener underflow solids	47 percent	49 percent

2. Aluminum. Concentrations to 200 mg/l were measured, presumably leached from ash penetrating the ESP, whereas the presence of aluminum was not detected with the fabric filter.
3. Fluoride. In contrast to relatively insignificant levels measured for the fabric filter; fluoride levels with the ESP/WS were as high as 500 to 700 ppm. Subsequent parallel tests with the fabric filter and ESP showed similar fluoride penetration from both devices. The reason that fluoride concentration increased to such a high level in the ESP/WS tests is not known. A synergistic effect of fluoride and aluminum promoting the high dissolved level of these constituents is suspected.
4. Total dissolved solids. Fifty percent greater with the ESP/WS due to high sulfate, magnesium, and sodium levels.
5. Calcium. Levels were only about half as much with the ESP/WS. The periodic high limestone levels were unable to increase available calcium levels.
6. Carbonate. Addition of tremendous quantities of excess carbonate (three times stoichiometric) did not effect pH. Carbonate levels were often as much as 60 percent of the solids in the scrubber slurry without raising the pH above 5.
7. Particulate Penetration. Particulate removal efficiency was just below 98 percent with the ESP compared with 99.99 percent with the fabric filter. These results were expected with the low SCA of 300 to 360.
8. Sulfur Dioxide Emissions. The SO₂ removal of the ESP/WS was limited compared to fabric filter tests as a result of ESP/WS lowered pH. SO₂ removal was limited to 69 percent with the ESP/WS compared with an average removal of 75 percent with no upper limit for the FF/WS if pH operating point was increased.

A review of the results indicated that aluminum and fluoride might be responsible for the limestone dissolution problem. A literature review revealed that tests completed by the Hitachi Research Laboratory (Reference No. 1) indicated that the dissolution of limestone is strongly affected by the presence of aluminum and fluoride. Bench-scale tests of the same phenomenon were completed by Brown and Caldwell. The results of these tests are presented in Figures 7 and 8.

The jar test results presented in Figure 7 were conducted as follows. Six flasks with 2.5 mls of deionized (DI) water were spiked with a standard solution of NaF, then varying amounts of AlCl₃ were added to achieve Al:F molar ratios ranging from zero (a seventh flask was used for control) to 5:1. Then 30 mls of a 15 percent CaCO₃ solution was added and the pH change

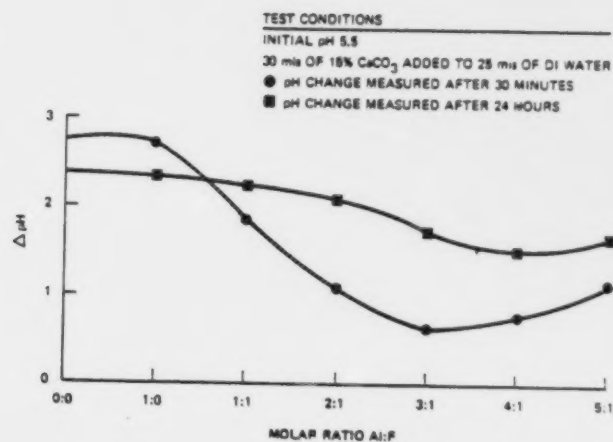


Figure 7 Effect of Aluminum and Fluoride on pH

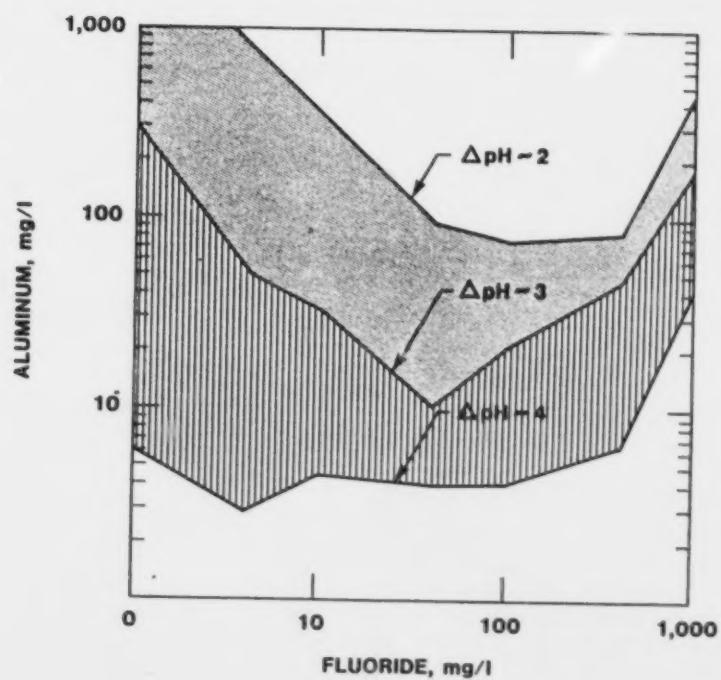


Figure 8 Change in pH of Limestone Solution with Varying Concentrations of Aluminum and Fluoride

was noted. The pH of the control increased by 2.8 pH units. The pH of the beaker with only aluminum present also changed 2.8 units. However, the pH change of the mixtures with 1:1, 2:1, and 3:1 molar ratios of Al:F showed lowered changes in pH as the Al:F ratio increased. The 3:1 solution registered a pH change of only 0.6 pH units. The 4:1 and 5:1 solutions registered slightly higher pH changes than the 3:1 but were still much lower than the control. The lowered pH was still present after 24 hours as shown in Figure 7. These results indicated to us that aluminum and fluoride did affect limestone dissolution. The change in pH over a 24-hour period indicated that the aluminum and fluoride probably affected the rate of limestone dissolution. Because the Al:F ratio during the ESP/WS tests varied from 1:4 to 1:6 a more complete matrix of the aluminum and fluoride concentrations were tested to develop a more complete picture of their impact on limestone dissolution. These results are presented in Figure 8.

The pH of the control test, with deionized water, increased 4.5 when limestone was added. The areas in Figure 8 indicate where the aluminum and fluoride restricted pH change to 4, 3 and 2 pH units. These bench-scale test results confirmed our suspicion that the aluminum and fluoride present in the scrubber slurry during the ESP/WS tests were responsible for the pH control problem and that aluminum and fluoride inhibit the dissolution of limestone.

These results suggested that more attention be paid to different particulate penetrations levels to the IECPP scrubber. Specific tests were then completed with the pilot plant to investigate the effect of particulate on the performance of the ESP/WS configuration. It was determined to run these tests at higher SO₂ levels to model the typical ESP/SO₂ operating conditions, i.e., higher sulfur coals.

High SO₂ Baseline Tests

These tests were conducted with an inlet SO₂ level of 2,000 ppm, closed loop water management, L/G ratio of 100, a limestone stoichiometric ratio of 1.05, and a desired pH control point of 5.4.

The results are summarized in Figure 9. The pH control problem was worse at high SO₂ test conditions than it had been at low SO₂ test conditions. The pH could not be raised above 4. Removal efficiency for SO₂ was 25 to 30 percent and the low pH of the scrubber slurry allowed significant off-gassing of SO₂ from the open reaction mix tanks. The aluminum levels were higher than under low SO₂ tests reflecting higher rates of aluminum dissolution from the fly ash at the lower operating pH levels. Fluoride levels were lower although not as low as expected from the increased purge rate at the higher SO₂ levels. The Al:F ratio was about 1:2.

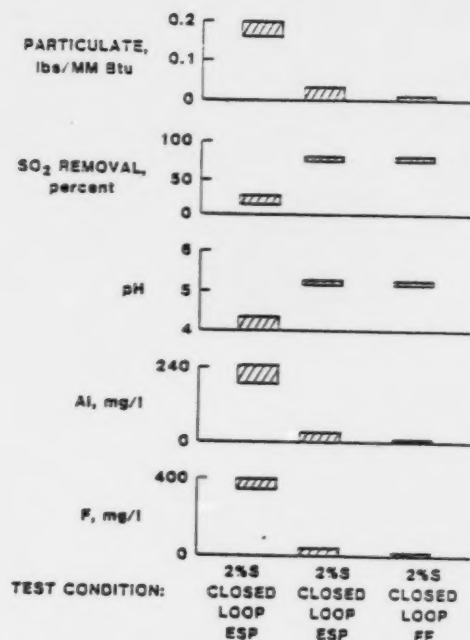


Figure 9 IECPP Test Results Effect of Particulate Penetration on SO₂ Control

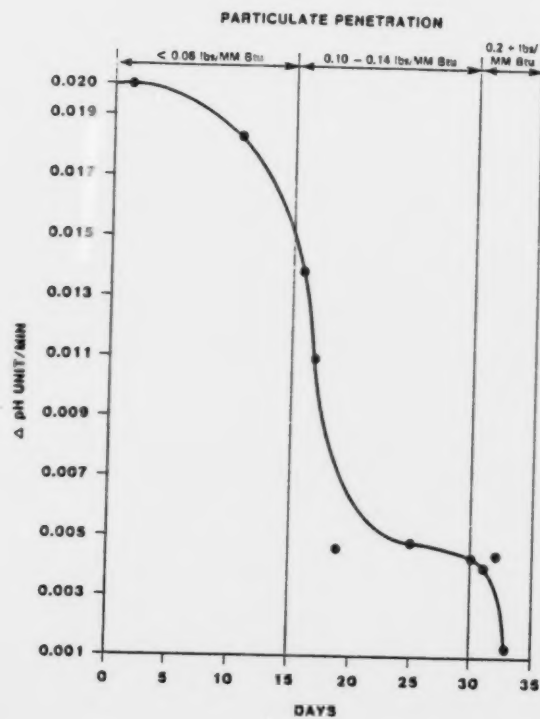


Figure 10 Effect of Particulate Loading on pH Response

Particulate Penetration

Since particulate levels into the scrubber were 200 times greater than with the FF/WS, the range of acceptable scrubber particulate loading was important in determining the extent of this problem in cases where retrofit SO₂ scrubbing may be required.

During high SO₂ tests, with particulate penetration levels between 0.15 and 0.2 lb/MMBtu, SO₂ removal was restricted below 30 percent and aluminum and fluoride levels remained high. These data are shown on Figure 9. When particulate penetration was lowered to 0.05 lbs/MMBtu, pH control, and SO₂ removal came into line with results from the FF/WS tests which are also presented in Figure 9. Thus, limestone dissolution interference does not appear to be a problem with newer ESP installations where particulate penetration levels should be low.

ESP Upset

One test was conducted to determine if the limestone dissolution problem could occur under a short-term upset ESP condition. Test conditions included pH at 5.4, limestone as the reagent and closed loop water management. Limestone dissolution and reactivity was measured by adding a known quantity of limestone to the reaction mix tank and recording the maximum change in pH over 10 minutes from this addition. This change in pH is plotted in Figure 10. At low particulate penetration levels (<0.06 lbs/MMBtu), this value was approximately ranged from 0.019 pH units/minute. At higher penetration levels (0.10 to 0.14 lb/MMBtu), this value dropped to 0.005 pH units/minute. The reader should note that it took approximately 3 days for the level to drop from 0.08 to 0.005. This is approximately three solids residence times for the IEC scrubber under the test condition. The penetration rate was then increased to 0.2 lb/MMBtu and again after 3 days the rate dropped further to 0.001 pH units/minute. Thus, although pH was always maintained at 5.4 during this period the limestone dissolution rate was minimal at the end of these tests. The aluminum and fluoride levels during these tests were quite low, generally less than 5 mg/l and 10 mg/l, respectively.

WATER MANAGEMENT

At higher inlet SO₂ levels, the scrubber solids production rate is higher than at low SO₂ levels and thus more water is purged from the system than at low inlet SO₂ levels. Thus, even under zero discharge operations, the concentration factor around the scrubber drops from 25 to 30 at low SO₂ (300-500 ppm) to 6 to 8 at 2,000 ppm. This reduction was expected to be sufficient to lower aluminum and fluoride concentration to a level where they did not affect limestone dissolution. As we saw under high SO₂ base-line tests, this did not occur. Under open loop water management conditions the concentration factor was lowered even further to 2. At this high water Purge rate, aluminum and fluoride levels were reduced to approximately

10 mg/l and pH control, at the desired operating point of 5.4, was achieved. The operating results are shown in Figure 11. These results showed that many ESP/WS which are presently operating may not be experiencing this problem because of their open loop water management operation.

The effect of opening the water loop around the scrubber is shown in Figure 12. The concentration of aluminum and fluoride was reduced by 1/3 from day 1 to day 2 and an additional 1/3 from day 2 to day 3. The concentrations then hovered near 50 mg/l and 100 mg/l for aluminum and fluoride for 2 days before dropping to less than 10 mg/l for both aluminum and fluoride on day 6. The system then showed excellent pH response and control. The reason for the sharp drop between days 5 and 6 is not known.

THIOSORBIC LIME TEST

Tests using thiosorbic lime were conducted to verify earlier batch tests that other reagents were not affected by the carryover of fly ash from the ESP. The results are summarized in Table 5 and are presented with results from similar limestone tests for comparison. The results for the open and closed loop lime tests are similar. The data also indicate 72 to 76 percent removal from the limestone tests. The reader should remember that in the case of the limestone tests where pH was limited to 5, the maximum SO₂ removal level achievable was 70 to 75 percent regardless of L/G rate or the amount of limestone added. The ultimate SO₂ removal was well over 95 percent with lime at much lower L/G ratios.

SOLID WASTE STUDIES

The solid waste from the ESP/WS tests, which were mixtures of ash and scrubber sludge in 1:1 and 2:1 ratios (wet), had significantly less strength and higher permeabilities than the samples from the FF/WS test. A summary comparison is presented in Table 6. The low SO₂ test results may be due to the higher oxidation levels of the FF/WS tests; however, the reason for the lower values during the high SO₂ tests are not clear. A complete matrix of solid waste mixtures will be completed which could indicate the reasons for the deterioration in waste solids characteristics.

SUMMARY AND APPLICATION OF RESULTS

The test results discussed in this paper indicate:

- Particulate penetration can complicate SO₂ removal when wet limestone scrubbers are operated under closed loop water management conditions. Particulate levels as low as 0.1 lb/MMBtu will effect limestone dissolution and thus scrubber performance depending on the aluminum and fluoride concentrations in the scrubber slurry.

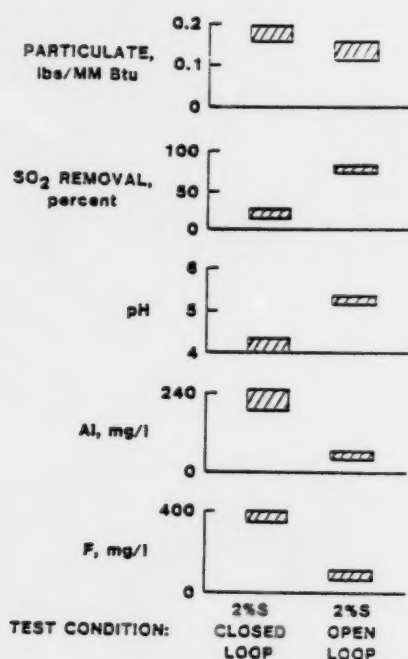


Figure 11 IECPP Test Results Effect of Water Management on SO₂ Control

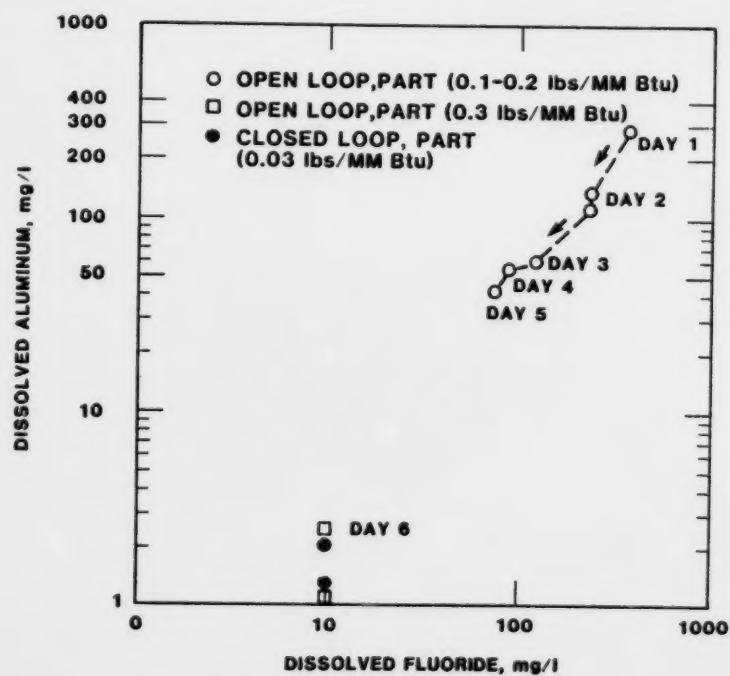


Figure 12 Change in Aluminum and Fluoride Concentrations Closed Loop to Open Loop Water Management System

Table 5. EFFECT OF THIOSORBICLIME AND LIMESTONE

	Lime	Lime	Limestone	Limestone
Test condition				
Percent S in fuel	2	2	2	2
Water management	Open	Closed	Open	Closed
Particulate penetration				
lbs/MMBtu	0.1	0.1	0.15-0.2	0.06-0.1
L/G	32	32	100	93
pH	6.2	6.2	5.0	5.5
Cycles of concentration	1.5	5.3	1.5	5.3
Test results				
SO ₂	75	75	72	76
Reagent stoich.	1.05	1.05	1.1	1.1

Table 6. COMPARISON OF SOLID WASTE RESULTS, LOW SO₂ TEST CONDITION

Test condition	Fly ash/filter cake/lime (percent in mixture)	Product moisture (percent)	Fabric filter/wet scrubber	ESP/wet scrubber	Fabric filter/wet scrubber	ESP/wet scrubber
Service water	45-50-5	22	395 ^a	192 ^b	10 ⁻⁷	7.26 x 10 ⁻³
Low SO ₂	62-33-5	22	1,010 ^a	N.M.	10 ⁻⁶	2.37 x 10 ⁻³
Service water	50-50-0	31	N.M. ^c	N.M. ^c	N.R.	4.43 x 10 ⁻³
High SO ₂	67-33-0	20	125 ^c	7 ^c	10 ⁻⁴	9.1 x 10 ⁻³

^aSludge oxidation 50 percent.^bSludge oxidation 100 percent.^cSludge oxidation 50 percent.Note: N.M. - not measurable.
N.R. - not recorded.

- System performance can be recovered by opening the water loop and purging accumulated dissolved levels of aluminum and fluoride; using a different reagent such as lime; or reducing particulate penetration. The water concentration factor may have to be reduced to as low as 1.5 to 2. Particulate penetration levels may have to be reduced to below 0.05 lb/MMBtu.
- The results of the solid waste tests indicate that a higher strength, lower permeability waste product is generated from the FF/WS than from the ESP/WS. The difference at low SO₂ was attributed to the higher oxidation level of the ESP/WS scrubber solids. However, the difference during high SO₂ tests is not clear.

The application of these results is appropriate for both new plants and retrofit limestone scrubbing systems for existing plants. A key concern is to minimize aluminum and fluoride carryover to a limestone scrubber. Potential sources of these compounds include makeup water, fly ash, gaseous fluorine, and limestone. If a retrofit limestone scrubber is being considered these gaseous components can be measured, if a new plant is being designed these constituents must be considered in engineering studies.

If significant levels of aluminum and fluoride are expected, the utility should consider (1) application for discharge of scrubber water and/or (2) the use of lime instead of limestone and/or (3) upgrading the particulate control device to minimize ash carryover.

REFERENCE

1. Effect of Al³⁺ and F⁻ on Desulfurization Reaction in the Limestone Scrubbing Process. Yoshikatsu Mori, Shimpel Matsuda, Furnitie Nakajma, Tsukawa Nishimura, and Yoshijiro Arikawa; Hitachi Research Laboratory Ind Eng Chem Process Des Dev, Vol. 20, No. 1, 1981.

KINETICS OF REACTION BETWEEN HYDRATED LIME AND SULFUR DIOXIDE

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ABSTRACT

The reactivity of dry hydrated lime with sulfur dioxide (SO_2) determines the extent of SO_2 removal by dust deposits in particulate control equipment following a spray dryer FGD system as well as by dried sorbent droplets in the spray dryer itself. In this work, the kinetics of this reaction was studied experimentally using a shallow bed of hydrated lime powder. Gas phase SO_2 concentration, gas phase relative humidity, and lime particle size were the principal variables investigated in this project. The relative humidity of the gas phase was found to have the most dominant effect on the nondimensionalized breakthrough curve of SO_2 concentration in the gas phase at the outlet of the bed. The reactivity of lime showed a weak dependence on the gas phase relative humidity; however, the lime utilization dramatically increased with increasing relative humidity of the gas phase. The reaction kinetics was of first order with respect to SO_2 concentration.

This paper has been reviewed in accordance with the U.S. Environmental Protection Agency's peer and administrative review policies and approved for presentation and publication.

INTRODUCTION

In a spray dryer flue gas desulfurization (FGD) system, particulate collection equipment (PCE) such as a baghouse or an electrostatic precipitator (ESP) is typically used following the spray dryer. The particulates deposited on the bags or on ESP plates often contain a significant amount of unreacted sorbent which may contribute to additional SO_2 removal. The SO_2 removal by such particulate deposits is also important in the E-SOX process and in the dry sorbent injection process. The reported SO_2 capture efficiencies by the particulate control equipment vary widely from less

than 5 percent to as much as 40 percent (e.g., Wilkinson and Tonn, 1982; Blythe and Rhudy, 1984; Gustke et al., 1984).

In general the SO₂ removal efficiencies in particulate collection equipment depend upon: 1) the stoichiometric ratio of sorbent in the primary process such as the spray dryer, as it determines the amount of available unreacted sorbent; and 2) the approach to saturation in both the spray dryer and the particulate collection device. The latter parameter determines the residual moisture content in solids which is thought to influence the reactivity of lime in the solid phase (Damle, 1986). A number of parameters (e.g., lime particle size, SO₂ concentration) are likely to affect the SO₂ removal process in the dry phase; however, no systematic data presently exist regarding the kinetics of the lime-SO₂ reaction at the typical conditions following a spray dryer. The reactivity of lime in the dry phase is also responsible for additional removal of SO₂ in a spray dryer after evaporation of sorbent droplets. In order to predict the extent of SO₂ removal in PCE, it is necessary to obtain the kinetic data for the lime-SO₂ reaction in the solid phase for a range of possible operating parameters. Such studies will also enable optimum utilization of the SO₂ removal potential of PCE.

In this project controlled experiments were carried out to determine the reactivity of lime in a shallow packed-bed reactor under a variety of conditions. The concentration of SO₂ in both the reactor inlet and the reactor outlet gas streams was monitored to determine the SO₂ concentration breakthrough curves for the lime bed. The breakthrough curves were analyzed to determine the reaction coefficients and the lime utilizations for various experimental conditions. The following sections describe the theory of noncatalyzed fluid-solid reaction, experimental apparatus, analysis methodology, and experimental results.

THEORY

Non-catalyzed fluid-solid reactions such as between SO₂ and hydrated lime may be described mathematically by an unreacted core model as shown in Figure 1. Since an intimate contact between sorbent particles and the gas phase is offered by a packed-bed reactor, the gas phase mass transfer processes for the reagent as well as product gas will not be controlling in determining the overall SO₂ removal. The SO₂ removal process may therefore be assumed to be controlled by the diffusion of SO₂ through the outer product layer and the chemical reaction at the outer surface of the unreacted core. For a single lime particle of radius R, with the unreacted core radius r_c, the molar rate of SO₂ removal, N (gmole/sec) is given by:

$$N = 4 \pi R^2 D \frac{dC}{dr} \bigg|_{r=R} = 4 \pi r_c^2 K C_s \quad (1)$$

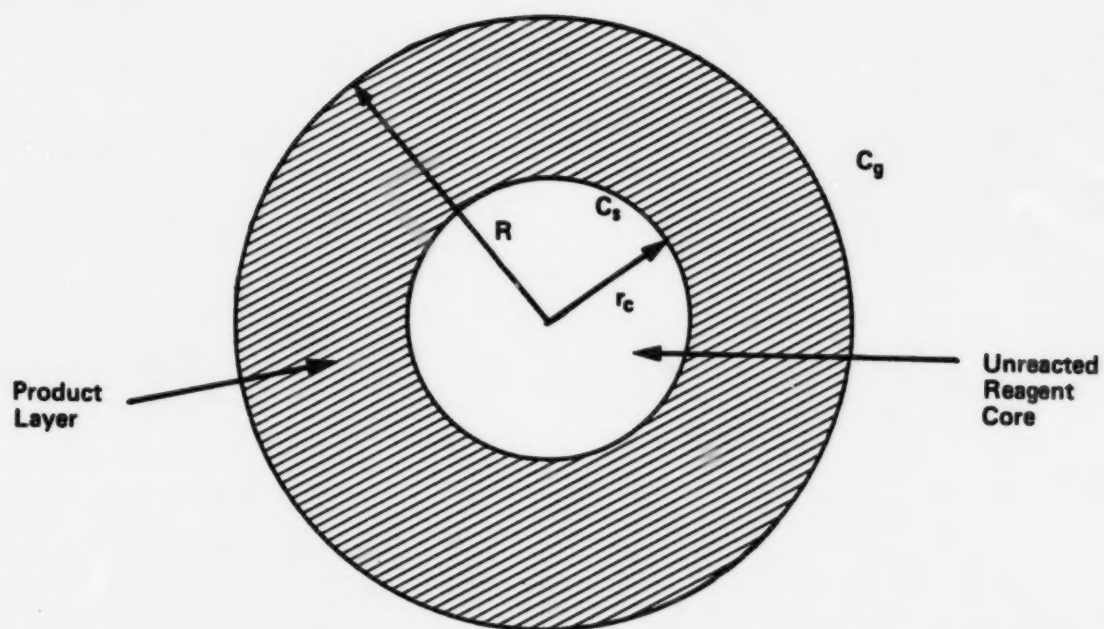


Figure 1. Schematic of unreacted shrinking core model.

where D is the diffusion coefficient, cm^2/sec

K is the reaction rate coefficient, cm/sec

C_g is gas phase SO_2 conc., gmole/cm^3

C_s is gas phase SO_2 conc. at the unreacted core surface, gmole/cm^3 .

The reaction rate equation assumes that the reaction rate is of first order with respect to SO_2 concentration. The surface concentration, C_s , may be eliminated from the above equation to obtain:

$$N = 4 \pi r_c^2 K C_g \frac{\phi}{1 + \phi} \quad (2)$$

$$\phi = \frac{D R}{K r_c (R - r_c)} \quad (3)$$

where, ϕ indicates the relative importance of the diffusion coefficient with respect to the reaction rate coefficient. For a small diffusion coefficient as compared to reaction rate, $\phi/(1+\phi) \rightarrow \phi$; whereas, for large D , $\phi/(1+\phi) \rightarrow 1$. The rate of SO_2 removal is maximum initially as $r_c = R$, and there is no diffusional resistance. As the product layer builds up, the rate of SO_2 removal will decrease depending upon ϕ . For a smaller ϕ , there will be a greater reduction in the rate of SO_2 removal.

The change in the unreacted core radius, r_c , with time due to reaction with SO_2 is given by mass balance as:

$$\frac{d r_c}{dt} = - \frac{K C_g M}{\rho} \frac{\phi}{1 + \phi} \quad (4)$$

where ρ is density, g/cm^3 , and M is the molecular weight of solid reagent.

The rate relations for a single particle may be integrated over the number of particles present in a section, dl , of a packed bed to obtain concentration profiles in a packed bed. The gas phase concentration and the unreacted core radius become functions of both time and axial position in the bed. The resulting equations may be written as:

$$\frac{d C_g(1,t)}{dl} = - \frac{4 \pi K r_c^2 N_p}{V} \frac{\phi}{1 + \phi} C_g(1,t) \quad (5)$$

$$\frac{d r_c(1,t)}{dt} = - \frac{K C_g(1,t) M}{\rho} \frac{\phi}{1 + \phi} \quad (6)$$

where, V = volumetric gas flow rate, cm^3/sec , and N_p = number of particles per unit depth of bed, number/cm.

At $t = 0$, $r_c = R$; and at $l = 0$, $C_g = C_{in}$.

With these boundary conditions, Equations (5) and (6) may be solved numerically to determine the SO_2 concentration profile in the packed bed at different times. Analytical solution is possible for the initial concentration profile ($t = 0$) as the unreacted core radius is constant for all particles in the bed and is equal to particle radius, R . Thus, the diffusion coefficient does not affect the rate of SO_2 removal at $t = 0$. The initial concentration of SO_2 at the bed outlet may thus be expressed as:

$$\frac{C_{out}}{C_{in}} = \exp - \frac{4 \pi R^2 N_p L K}{V} \quad (7)$$

where L is the depth of the packed bed.

It is interesting to note that, for first order kinetics, the concentration profile or the ratio of the outlet and inlet concentrations is independent of the inlet gas phase SO_2 concentration. By rearranging, the reaction rate K may be obtained as:

$$K = - \frac{V}{4 \pi R^2 N_p L} \ln \frac{C_{out}}{C_{in}} \quad (8)$$

The reaction rate coefficient K as defined by Equation (1) is based on the available surface area of a lime particle. A more convenient definition may be based on the total mass of the sorbent in the bed, which will then describe the overall reaction rate coefficient. The overall reaction rate coefficient K' , $\text{cm}^3/\text{sec-g}$, is related to K by:

$$K' = \frac{3 K}{R \rho} = - \frac{V}{W} \ln \frac{C_{out}}{C_{in}} \quad (9)$$

where W is the weight of sorbent in the bed.

The above definition of the rate coefficient may also be applied for any given time t , to obtain the effective reaction rate coefficient of the partially reacted lime. This effective coefficient will include the diffusional resistance of the product layer to the overall SO_2 removal process. This concept of effective reaction rate coefficient will allow direct utilization of the experimental data to predict SO_2 removal in the particulate control equipment.

EXPERIMENTAL

APPARATUS

The dust deposited on filter bags or electrostatic precipitator plates was simulated in these experiments by a shallow packed-bed of lime powder. Figure 2 shows the schematic of the experimental apparatus used. As shown in this figure, a series of two bubblers immersed in a temperature-controlled water bath was used to humidify compressed air. A filtered compressed air stream was split in two parts: one part of the air was passed through the bubblers while the other part bypassed the bubblers and was then remixed with the humidified part of air. By controlling the ratio of the bypass air flow rate to the flow rate of air passing through the bubblers, the humidity of the total gas stream could be varied. The humidification efficiency of the bubblers at various flow rates was determined prior to the reactivity experiments and was found to be within 70 to 75 percent. These calibrations were used, along with the respective air flow rates through the bubblers and bypass stream, to determine the humidity of the total gas stream in the kinetic experiments. The total compressed air flow rate was controlled by a mass flow controller.

The humidified air was mixed with 1 percent SO_2 in nitrogen to obtain various levels of SO_2 concentration. By controlling the SO_2 stream flow rate by a mass flow controller, desired SO_2 concentration levels were obtained. A 3 in. (7.6 cm)-I.D. glass pipe was used as the packed-bed reactor. The shallow bed of lime was supported by a fritted glass support and filter paper. The temperature of the reactor and the bed was controlled by a heating tape wound around the reactor. Stainless steel tubing was used for all the gas lines, and all the tubing sections carrying humidified air were heated to prevent condensation of water vapor. The air/ SO_2 stream was passed downward through the bed. The bed depth was typically 3 to 6 mm.

The air streams at the inlet of the bed and at the outlet of the bed were sampled through a 4-way switching valve. The purpose of using a 4-way valve was to draw a constant sample flow of 1 lpm of air from both the inlet and the outlet gas streams at all times. Of the two streams leading from the 4-way valve, one was sent to a DuPont 411 photometric analyzer while the other was bypassed as shown in Figure 1. The flow rate in each sampled stream was controlled by critical orifices. This sampling arrangement allowed switching between the inlet and outlet gas streams for SO_2 concentration monitoring without affecting the flow through the packed bed. The SO_2 concentration was measured by the DuPont 411 photometric analyzer which measures attenuation of ultraviolet light by SO_2 in the gas.

VARIABLES STUDIED

The primary variables studied in this experimental program were the gas phase relative humidity and the SO_2 concentration in the gas phase. Four levels of gas phase relative humidities (0, 25, 50, and 75 percent) were used, and the SO_2 concentration was varied between 300 and 550 ppm.

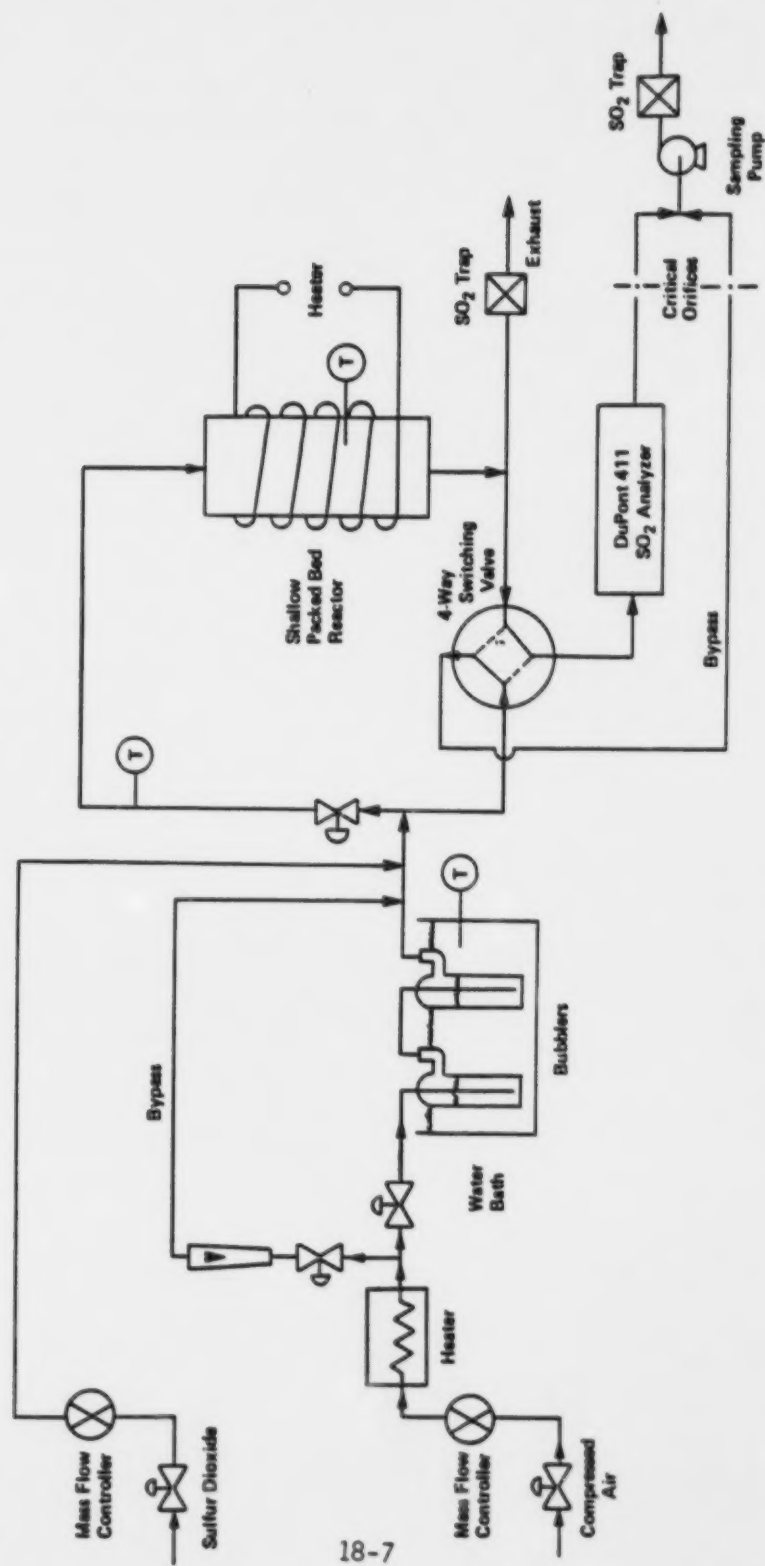


Figure 2. Schematic of experimental system.

Fine grade, high-calcium Tenn-Luttrell hydrated lime was used in these experiments. The mean lime particle diameter was about 10 μm . The product solids from a typical spray dryer FGD plant are usually much coarser. Therefore, to study the effect of lime particle size on SO_2 removal, a small amount of the above mentioned fine lime was slurried and spray dried at the U.S. EPA's pilot scale E-SOX and precipitator facility in the Research Triangle Park. This spray-dried lime had a mean particle diameter of 38 μm . Experiments were also carried out with this batch of lime, referred to as coarse lime. Both of these batches were pure lime. The spray dryer FGD product solids contain unreacted lime and lime- SO_2 reaction products in different proportions depending upon the stoichiometric ratio and SO_2 removal efficiency in the spray dryer. In order to assess the effect of product solids composition, small amounts of product solids were obtained from the spray dryer FGD pilot plant at the University of Tennessee. A few experiments were also carried out with these product solids samples.

The packed-bed reactor experiments were run at gas phase conditions typical of those in a baghouse following a spray dryer FGD plant. The temperature of the reactor was maintained at 70 $^{\circ}\text{C}$, and the gas flow rate through the packed bed was 2.8 lpm with the corresponding flow velocity of 1 cm/sec. Typical gas phase relative humidities following a spray dryer FGD unit range from 20 to 60 percent, depending upon the approach to saturation in the spray dryer; whereas, the experiments covered humidities from 0 to 75 percent. For dry runs (0 percent humidity), 10 g of lime was used in the bed. The corresponding bed depth was found to be approximately 6 mm. The breakthrough time for these dry runs was found to range from 20 to 90 minutes depending upon the lime used and the concentration of SO_2 . For humid conditions the breakthrough time was found to increase dramatically. Therefore, for humid runs, the amount of lime used was reduced by half (to 5 g) to limit the breakthrough times. The corresponding bed depth was 3 mm. Typical dust cake thickness in a baghouse is about 1 cm. Thus, the experimental bed depth was somewhat shallower than a typical filter cake thickness.

DATA ANALYSIS METHODOLOGY

A typical breakthrough curve (i.e., a plot of nondimensionalized SO_2 concentration at the bed outlet with respect to exposure time) obtained in these kinetics experiments is shown in Figure 3. Such a breakthrough curve was observed in all experiments except those carried out with 75 percent humidity conditions. In the high humidity runs, a steady state outlet concentration was observed for a long period of time before the run was terminated. The breakthrough curve shown in Figure 3 indicates the initial outlet SO_2 concentration, the breakthrough time, and the total integrated amount of SO_2 removed by the bed and hence the utilization of lime in the bed.

The outlet SO_2 concentration in all breakthrough curves initially increases rapidly without significant consumption of lime. This indicates

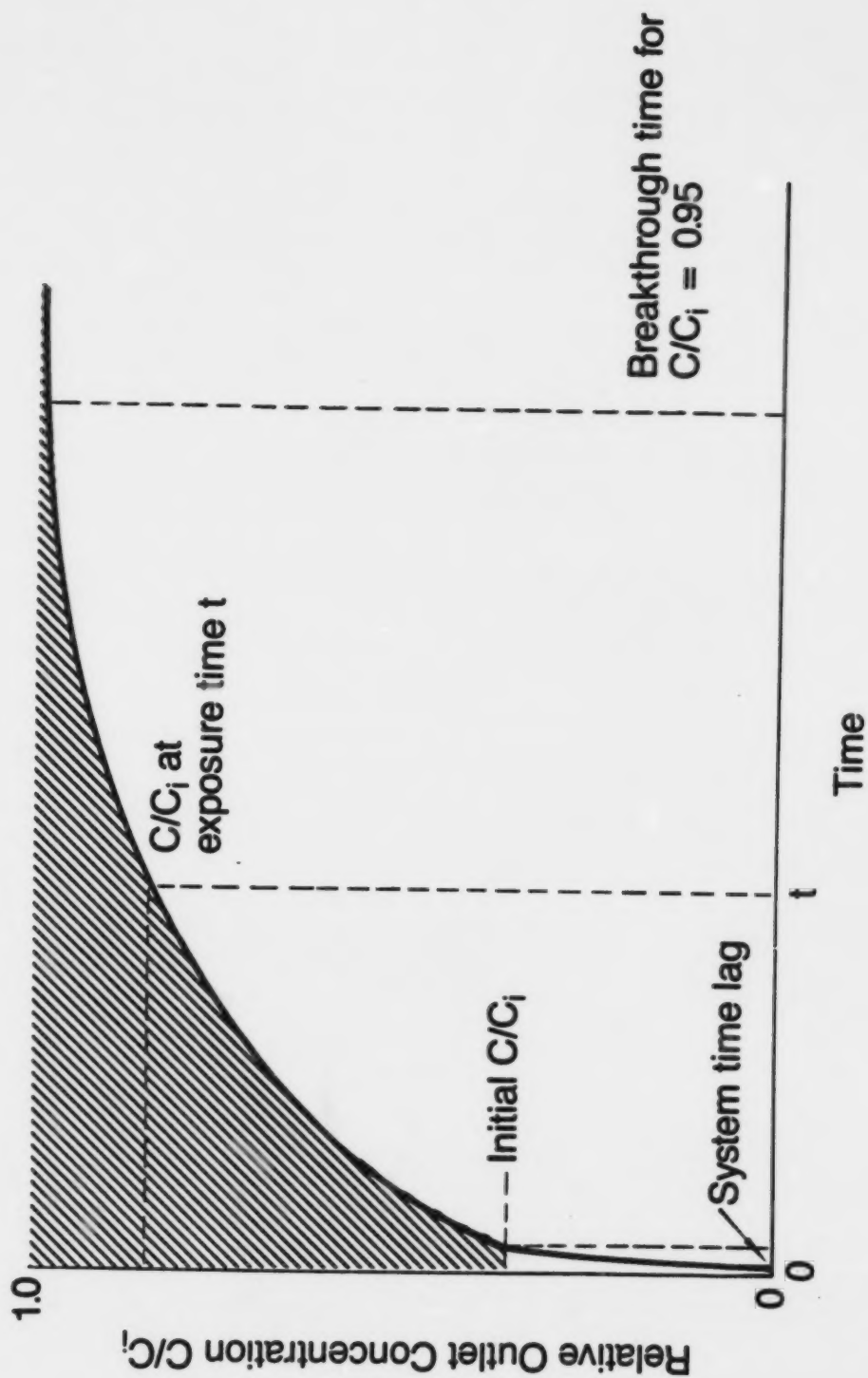


Figure 3. A typical breakthrough curve.

a strong influence of diffusion of SO_2 and a low SO_2 diffusivity in the outer product layer. Another important observation was that the initial outlet nondimensionalized SO_2 concentration for a given lime and gas phase humidity was found to be independent of the inlet SO_2 concentration. As discussed in the theory section [Equation (7)], this fact indicates a first order kinetics with respect to SO_2 concentration. The reaction rate coefficient of pure sorbent, K , may thus be determined directly from the initial outlet concentration using Equation (8) or (9). The observed breakthrough curve may then be compared to that predicted by solution of Equations (5) and (6), to obtain the SO_2 diffusion coefficient, D , in the product layer.

The outlet concentration at any exposure time, t , may be used to obtain the effective reaction rate coefficient, K_{eff} , by using Equation (9). The effective rate coefficient includes the effect of diffusion in the product layer. Initially, since no diffusional resistance is present, the same equation determines the reaction rate coefficient of pure lime. The effective reaction rate coefficient at a given exposure time, t , can be used directly to estimate SO_2 removal in particulate control equipment, as discussed later.

The experimental data obtained under various conditions were thus analyzed to determine the initial outlet SO_2 concentration, the corresponding rate coefficient K , the breakthrough time, and the lime utilization during the breakthrough period. The breakthrough curves were also compared with the theoretical predictions to determine the SO_2 diffusion coefficient. The effect of gas phase relative humidity, lime used, and the gas phase concentration on the reaction rate coefficient, lime utilization, and breakthrough time was determined.

EXPERIMENTAL RESULTS

The experimental results obtained with pure fine lime are summarized in Table 1. As seen from this table, the gas phase relative humidity has a moderate influence on the reaction rate coefficient, K , and a strong influence on the percent lime utilization and the breakthrough time. The gas phase SO_2 concentration has no significant effect on the reaction rate coefficient and the percent lime utilization. Since the percent lime utilization is independent of SO_2 concentration, it follows that the breakthrough time and the shape of the breakthrough curve would be dependent on the SO_2 concentration. For lower inlet SO_2 concentration, the breakthrough time is greater to match the same lime utilization. The breakthrough curve was often not complete, especially for 50 percent humidity runs, during an experimental run, in which case the tailing portion of the breakthrough curve was extrapolated. For 75 percent humidity runs the outlet concentrations were in steady state for a long period of time until the runs were terminated. In these cases the breakthrough curve was not extrapolated and neither the final lime utilization nor the breakthrough time could be determined. The results indicate that the initial lime reaction rate coefficient increased by almost a factor of 4 by increasing the humidity

TABLE 1. EXPERIMENTAL RESULTS: FINE HYDRATED LIME—MEAN SIZE ~ 10 μ m

	% Relative Humidity			
	0	25	50	75
Rate coefficient K ($\text{cm}^3/\text{sec-g}$)	6.7	16.2	23.3	22.7
Lime utilization (%)	1.7	6.6	25.8	> 40
Breakthrough time (min)	75-90	200-400	600-1,000	> 1,000

from 0 to 50 percent; whereas, the lime utilization increased by more than an order of magnitude by the same increase in the humidity.

The lime utilization during the breakthrough period was only 1.7 percent for the dry conditions. This indicates a diffusion-controlled process under dry conditions. Comparison of the breakthrough curves for the dry conditions with the theoretical predictions gave a good match when the diffusivity value was 1.0×10^{-9} cm²/sec. The theoretical predictions, however, did not match with the experimental breakthrough curves for 25, 50, and 75 percent humidity runs. Thus, a simple shrinking core model did not appear to be applicable for runs with gas phase humidity. Increased lime utilization with increasing humidity suggests formation of a moisture film on the reagent particles which may prevent buildup of a uniform product layer and make more reagent available for reaction. The steady-state observation for 75 percent humidity runs may possibly be explained by the presence of a persistent moisture film and continued steady-state dissolution of lime.

The experimental results obtained with pure coarse lime are summarized in Table 2. The results obtained with coarse lime follow similar trends as seen for fine lime. The reaction rate coefficient increased moderately with increasing relative humidity; whereas, the lime utilization and breakthrough time increased dramatically with humidity. Here also for 75 percent humidity runs, a steady state was observed for a long period of time until the runs were terminated.

The reaction rate coefficient for coarse lime under dry conditions was found to be similar to that for fine lime. The reaction coefficient increased at higher humidities although the increase was somewhat lower as compared to fine lime. At 75 percent humidity, the rate coefficients were again comparable. The lime utilization and the breakthrough times were consistently lower than those for fine lime under similar conditions. The reaction rate coefficient and the lime utilization were again found to be independent of the inlet SO₂ concentration. The breakthrough time, therefore, was dependent upon the inlet SO₂ concentration. The lower utilization and somewhat lower reactivity may be attributed to lower surface area to volume ratio for the coarser lime. The mean size of the coarse lime particles was almost three times that of the fine lime mean particle size. Consequently, the surface area to volume ratio for coarse lime was smaller by a factor of three as compared to fine lime. The lime utilization was also found to be lower for coarse lime by a factor of two to three under similar conditions. Thus, lime particle size appears to influence the lime utilization in inverse proportion.

The results obtained with spray dryer FGD product samples are summarized in Table 3. Runs were made only with dry and 50 percent relative humidity conditions. The observed reaction rate coefficients and lime utilizations compare well with pure lime results after taking into account the mass fraction of unreacted lime in spray dryer FGD solids which was about 40 to 50 percent. This comparison is only qualitative, as the sources of the pure lime used in the present study and that for the lime in

TABLE 2. EXPERIMENTAL RESULTS: SPRAY DRIED LIME—MEAN SIZE ~ 35 μm

	% Relative Humidity			
	0	25	50	75
Rate coefficient K ($\text{cm}^3/\text{sec-g}$)	6.7	11.1	12.4	23.9
Lime utilization (%)	0.5	3.5	13.3	> 40
Breakthrough time (min)	20-30	180-280	250-600	> 1,000

TABLE 3. EXPERIMENTAL RESULTS: SPRAY DRYER FGD PRODUCT SOLIDS

	% Relative Humidity	
	0	50
Rate coefficient K (cm ³ /sec-g)	3.6	8.9
Lime utilization (%)	1.0	15.0
Breakthrough time (min)	80-120	500-700

spray dryer FGD pilot plant were not the same. The present results qualitatively indicate that the reactivity and utilization of spray dryer product solids may be estimated by using the corresponding values for pure lime and multiplying them by the mass fraction of unreacted sorbent.

One interesting observation is that the spray dryer FGD solids behaved in a similar way as pure lime in spite of the unreacted lime content being less than 50 percent of the total mass. Breakthrough or deactivation of these solids was observed with only 1 percent of additional lime utilization under dry conditions and about 15 percent of additional lime utilization in the case of 50 percent relative humidity conditions. This indicates that different mechanisms are at work for SO_2 removal during drying of a slurry droplet in spray dryer FGD and in the dry phase in the packed bed. The deactivation or the breakthrough during packed bed experiments is thought to be due to the buildup of product layer on the reagent particles. Since the spray dryer FGD solids were still active even after consumption of 50 percent of the initial lime reagent, the SO_2 removal during drying of slurry droplets was not accompanied by a uniform product layer buildup. Also, the fact that the reactivity of spray dryer FGD product solids could be related to the mass fraction of unreacted reagent indicates that the reaction products and the unreacted sorbent were uniformly distributed throughout the dried slurry droplets due to a uniform aqueous phase reaction.

SO_2 REMOVAL IN PARTICULATE CONTROL EQUIPMENT

The effective reaction rate coefficient, K_{eff} , as determined in the packed bed experiments for certain conditions, can be used to determine the extent of SO_2 removal in particulate control equipment following the spray dryer under similar conditions. In the spray dryer, flue gas contacts only the sorbent fed into the spray dryer; whereas, in a particulate collection device, the gas phase may contact solids accumulated over a long length of time. This effectively tends to provide a very high stoichiometric ratio in the particulate collection device.

As seen from the breakthrough curves, the SO_2 removal efficiency of given solids depends upon its exposure time or, more specifically, the amount of SO_2 that a unit mass of the sorbent has been exposed to. The dust deposited on fabric filter bags and precipitator plates is usually subjected to a certain cleaning schedule. As a result, in a steady state, the dust deposits will have a certain distribution of exposure time. If this exposure time distribution or a mean dust exposure time is known, the amount of SO_2 that an average mass of the dust is exposed to can be determined. Using the corresponding exposure time on the lime breakthrough curve under similar conditions of the dust and dust environment, the effective reaction rate coefficient applicable in the actual situation can be determined. The extent of SO_2 removal in particulate control equipment can then be determined by appropriate mass balance equations derived for a baghouse and an ESP in the following sections.

BAGHOUSE

As the flue gas passes through the dust cake, SO_2 is transported from the bulk-gas phase to the dust surface. Due to intimate dust-gas contact, the SO_2 concentration at the dust surface will be the same as the bulk gas phase SO_2 concentration (Figure 4). The efficiency of SO_2 removal, η , in the dust cake is then given by rearranging Equation (9) derived earlier for a packed bed of lime particles:

$$\eta = \frac{C_{\text{in}} - C_{\text{out}}}{C_{\text{in}}} = 1 - \exp - \frac{K_{\text{eff}} W}{V} \quad (9')$$

where, W is the amount of unreacted sorbent in the dust cake per unit surface area of a bag, V is the volumetric gas flow rate through a unit surface area of a bag, and K_{eff} , $\text{cm}^3/\text{sec-g}$, is the effective overall reaction rate coefficient for the conditions prevailing at the dust cake.

Thus, the efficiency of SO_2 removal in the baghouse is expected to increase with higher reactivity (K_{eff}), a greater amount of unreacted sorbent in the dust cake, and a lower air-to-cloth ratio.

ELECTROSTATIC PRECIPITATOR

In an ESP, gas phase diffusion transports SO_2 from the bulk gas phase to the dust and unreacted sorbent is deposited on the collection plates. Therefore, the gas phase diffusion resistance contributes to overall SO_2 removal. A schematic of an ESP section is shown in Figure 5. By taking a mass balance on the small section shown in this figure, the rate of SO_2 removal may be given as:

$$-V dC_g = K_d (C_g - C_s) dA = K_{\text{eff}} C_s W dA \quad (10)$$

where K_d is the gas phase mass transfer coefficient, C_g is the bulk gas phase SO_2 concentration, C_s is the concentration of SO_2 at the dust surface, A is the surface area of the collection plates, K_{eff} is the effective reaction rate coefficient for the conditions prevailing at the dust cake, and W is the amount of unreacted sorbent per unit surface area of the collection plates. Eliminating surface SO_2 concentration, C_s , the change in the SO_2 concentration as the flue gas flows over a section dA may be obtained as:

$$-V \frac{dC_g}{dA} = K_d C_g \left(1 - \frac{K_d}{K_d + K_{\text{eff}} W} \right) \quad (11)$$

By integrating Equation (11) over the entire ESP surface area, total SO_2 removal efficiency of the ESP can be determined. Both the effective reaction rate coefficient and the gas phase mass transfer coefficient may be important in determining overall SO_2 removal. ESPs have very well-defined flow geometry (flow through two parallel plates) and, hence, experimental mass transfer correlations reported in the literature for

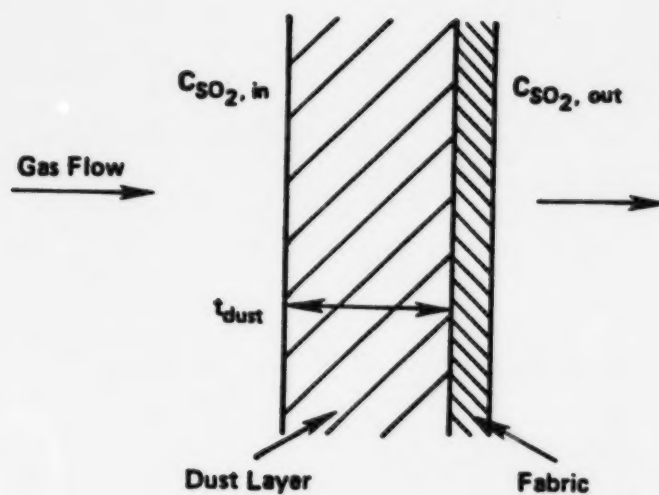


Figure 4. Schematic of SO_2 removal in a baghouse.

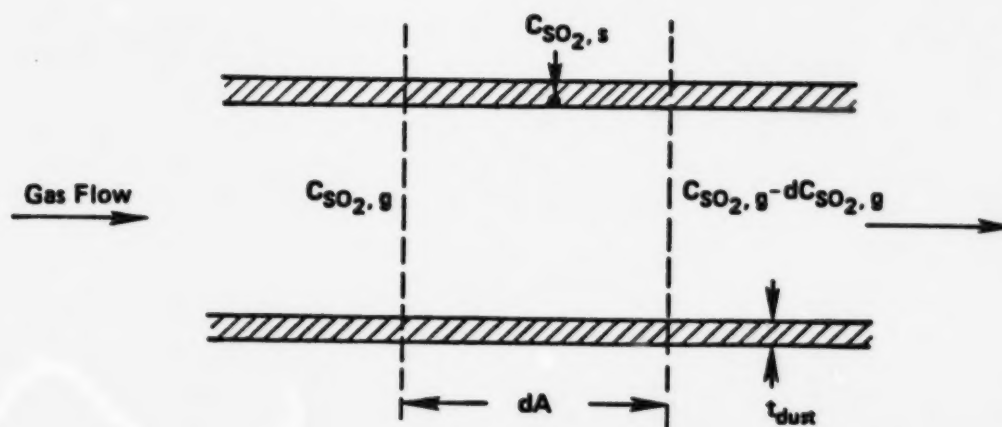


Figure 5. Schematic of SO_2 removal in an electrostatic precipitator.

similar flow geometries may be used to estimate the mass transfer coefficient, K_d . For a high mass transfer coefficient compared to reactivity, the dust surface concentration of SO_2 will approach that in the bulk gas phase. Conversely, for a large reaction rate coefficient, the surface concentration will approach zero.

SO_2 REMOVAL IN DRY PHASE IN SPRAY DRYER

Since the residence time of spray droplets in a spray dryer is of the order of only 10 sec, the initial reaction rate coefficient of the spray dryer solids may be used to determine the additional SO_2 removal by dried spray droplets. For example, consider a spray dryer FGD system with the following operating parameters:

Stoichiometric ratio - 1.0
 Inlet SO_2 concentration - 1000 ppm
 No recycle
 Exit temperature - 150 °F (65.6 °C)
 Exit relative humidity - 50%
 (approach to saturation - 24 °F (13.3 °C))
 Total residence time - 12 sec
 Residence time (dry phase) - 10 sec
 Efficiency of SO_2 removal in wet phase - 60%

The SO_2 concentration at the spray dryer exit would be 400 ppm without any additional SO_2 removal in the dry phase. The reaction rate coefficient of the spray dryer solids, as determined in the present experiments under conditions similar to the above mentioned operating conditions, is expected to be 8.9 cm³/sec-g of sorbent. With 60 percent of lime utilization in the wet phase, the remaining lime in the dried droplets will be $0.4 \times 10^{-3} \times 74$ g/gmole of gas or 1.066×10^{-6} g/cm³ of gas. The reduction in SO_2 concentration in the dry phase in ppm is then calculated as:

$$dC = - \frac{K_{eff} W}{V} C dt$$

$$dC = - 8.9 \times 1.066 \times 10^{-6} \times 400 \times 10.$$

$$(ppm) (cm^3/sec-g) (g/cm^3) (ppm) (sec)$$

$$= - 0.038 ppm.$$

Thus, the extent of SO_2 removal by dried droplets in a spray dryer is expected to be negligible based on the experimental reaction rate coefficient determined in this study.

CONCLUSIONS

The following conclusions may be drawn from the experimental data obtained in this project:

1. Among all experimental parameters, the gas phase relative humidity was found to have the most influence on the reaction kinetics for the reaction between dry hydrated lime and SO_2 .
2. The reaction rate coefficient increases moderately (by a factor of three to four) with increasing relative humidity from 0 to 50 percent. The lime utilization and the breakthrough time, however, increase dramatically (more than an order of magnitude) with the same increase in relative humidity.
3. The experimental runs at 0, 25, and 50 percent relative humidities resulted in breakthrough curves with continuously increasing outlet concentrations. The experimental runs at 75 percent humidity resulted in a steady-state outlet concentration until the runs were terminated.
4. Within the range studied, SO_2 concentration was found to have no effect on the reaction rate coefficient and the lime utilization. The breakthrough time was found to depend on the concentration. Lower concentrations required longer breakthrough times to achieve the same utilization.
5. Fine lime was found to have greater utilization and somewhat better reactivity as compared to spray-dried coarse lime.
6. Spray dryer product solids' reactivity and utilization are related to those of pure lime by the mass fraction of unreacted lime in solids.
7. The "Shrinking Core" model fits data well only for dry conditions, with the effective pore diffusivity of SO_2 in the product layer about $10^{-9} \text{ cm}^2/\text{sec}$.
8. The concept of the effective reaction rate coefficient, K_{eff} , at a certain exposure time can be used to determine the extent of SO_2 removal in particulate control equipment.
9. Based on the experimental data, the extent of SO_2 removal by dried droplets in spray dryer FGD is negligible.

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Mini Spray Dryers for Low-Cost SO₂ Control

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ABSTRACT

Potential retrofit SO₂ control technologies have been reviewed and assessed conceptually, based upon very limited pilot plant experience. Of the candidate retrofit control technologies for SO₂ control, a small spray dryer, coupled with an existing electrostatic precipitator, offers promise for a cost-effective combined control system. A small or mini spray dryer operates with a 2-3 sec residence time instead of 10 sec for conventional spray drying but has considerably less risk of failure than proposed in-duct spray drying schemes. At modest calcium to sulfur ratios of 1:1, 50 percent SO₂ removal is anticipated. Higher SO₂ removals of 70 percent or greater using advanced sorbents and other process concepts appear achievable.

Current research at EPA facilities in the Research Triangle Park indicates that the mini spray dryer is a viable technology for applications in which moderate SO₂ removal might be required, such as potential acid deposition mitigation. It will continue to focus on examination of fundamental SO₂ removal mechanisms and characterization of process options, including sorbents, mass transfer additives, slurry recycle, duct conditions, and upgrading of electrostatic precipitators to optimize the overall control of both SO₂ and particulate matter.

Preliminary capital cost estimates for 200 and 40 MW_e retrofits show significant advantages over conventional SO₂ control retrofit.

Introduction

The history of development of flue gas desulfurization (FGD) technology has been one of trial and error where the driving forces (emission regulations) preceded a fundamental understanding of system chemistry, and economics (cost-effectiveness) were difficult to predict. Now that technology has matured, a new driving force on the horizon will cause a reevaluation of alternative FGD technologies and more focus on the underlying process economics (and, hence, on fundamental process chemistry).

This new driving force is the perceived SO₂ emissions reductions which are likely to be mandated by forthcoming National and/or Regional acid rain abatement strategies, new Federal Tall Stack Regulations, and subsequent revisions to State Implementation Plans as a result of these Federal actions. The targets are existing non-FGD utility and large industrial boilers, most of which are currently equipped with electrostatic precipitators (ESPs). With the anticipated SO₂ removal requirements being somewhat less than the 70-90 percent level currently required by NSPS, and the potentially shorter life of the affected facility, focus is duly given to systems which have lower capital costs, even if operating costs (most notably reagent costs) are higher.

The SO₂ reduction "arena" is once again open to a number of precombustion, combustion, and post-combustion technologies with promises of moderate SO₂ removal at lower capital cost. Low capital cost generally precludes the higher technology precombustion techniques such as chemical coal cleaning, liquefaction, and other synfuels variations. Even physical coal cleaning may require substantial capital investment by the user, the Homer City facility being a \$120 million example. Similarly, conventional flue gas desulfurization, whether wet or dry, incurs substantial capital investment (and long-term commitment to operate). Therefore, current focus is upon in-furnace and in-duct injection of SO₂ sorbents and their collection with existing particulate matter controls. The remainder of this discussion is devoted to duct injection, although potential improvement of in-furnace concepts is implicit in the following discussions of bench-scale work.

The authors' opinions expressed herein are based upon a review of currently proposed concepts and upon our ongoing research with in-house pilot plants, the recent work in E-SOX, and reported results from Conoco's field demonstration and pilot plant work.

Because of the issues of solid waste handling and disposal which arise from the use of sodium sorbents, the discussion will focus upon calcium slurries and dry calcium powders as the SO₂ sorbent.

Current Proposed Schemes for Retrofit SO₂ Control

Several recent schemes for in-duct retrofit SO₂ control, including "coolside" furnace injection enhancement, have been presented. A select few are reviewed herein to focus later discussion on needed areas of research; a comprehensive review of the myriad concepts currently under consideration is intentionally avoided for this reason.

The most simple concept consists of spraying a reactive slurry in existing ductwork such that the slurry dries before wetting the walls or contacting downstream surfaces, such as the internals of an electrostatic precipitator. An increase in flue gas solids loading may be offset by cooling of gases which increases the gas residence time within, and hence the collection efficiency of, the precipitator. One noted example of this concept is the "Confined Zone Dispersion" (CZD) process currently advanced by Bechtel Incorporated.¹ Other than a slurry preparation and feed system, the only additional hardware required with CZD is an injection and dispersion apparatus, currently projected to consist of a single dual-fluid spray nozzle, using compressed air. Pressure hydrated dolomitic lime is indicated as the preferred slurry material, based upon bench-scale and limited pilot plant evaluation, using a spray dryer and fabric filter combination. Process requirements include a 50 to 100 ft (15-30 m) length of straight duct downstream of the spray (1 to 2 sec gas residence time), reasonably uniform flow, and no internal structures where impingement of droplets may occur.

Bechtel estimates that at least 50 percent capture of SO₂ with high lime utilization is indicated with this arrangement.

General Electric (GE) has advanced a similar concept to that of Bechtel, except that a rotary atomizer is proposed for slurry injection.² GE's choice of a rotary atomizer is based upon the flexibility of changing slurry feed rates without changing the droplet size distribution. System performance of 50 percent SO₂ removal with 1 sec duct residence time using lime slurry reagent is indicated.

Conoco has advanced further than the previous developments in what is known as "Coolside Desulfurization" by having performed field trials on a 3000-4000 acfm (85-113 m³/min) slipstream from a small pulverized coal boiler.³ The Conoco data suggest that (1) dry calcium hydroxide is not active enough to obtain high SO₂ removals, (2) wetting of the injected solids is important in sorbent activation, (3) additives such as sodium hydroxide are necessary to achieve 50 percent SO₂ removal in the duct, and (4) approach to saturation must be in the 20-25°F (11-14°C) range to achieve greater than 50 percent SO₂ removal. Conoco devoted appreciable laboratory (bench-scale) reactor efforts in determining the sorbents to be field tested.

One other effort to be mentioned is the development of E-SOX by EPA at the Research Triangle Park.⁴ This concept is based upon modifying an existing ESP to a combination spray dryer and advanced ESP. By upgrading the downstream ESP sections, particulate matter control may be maintained to original specifications while the first ESP section is equipped with a slurry injection system, becoming a spray dryer chamber of 1.5 to 3 sec gas residence time.

Results from the E-SOX in-house pilot plant indicate that 60 percent SO₂ removal with E-SOX is possible with a residence time of 1 sec using hydrated lime slurry as the sorbent at moderate (1.3-1.4) reagent/SO₂ molar ratios. Higher SO₂ removals (up to 90 percent) were obtained using sodium carbonate solution with E-SOX.

Assessment of In-Duct Spray Drying to Date

We are not aware of any successful application of true in-duct spray drying to date, on a commercial or pilot basis. While the economics of such concepts are compelling, their ability to get good mass transfer or mixing of sorbent with flue gas in a short residence time without wetting the duct walls appears risky. The only reportedly successful pilot concept (Conoco) used a duct expansion/humidifier, mass transfer additive (Na₂CO₃), a relatively high sorbent/SO₂ ratio (2:1), and a 20-25°F (11-14°C) approach to saturation.

Based upon the Conoco experience and our own pilot plant experience at RTP, we feel that a type of duct expansion with water/slurry spray drying offers the best compromise between low capital cost and SO₂ removal effectiveness. Several of the referenced processes previously discussed emphasize that only 2-3 sec drying time is needed for slurry droplets in a duct where flue gas temperatures are on the order of 150°F (83°C) above saturation. Moreover, based on a reported survey by General Electric, over 90 percent of existing utility systems would have the ductwork length which could accommodate a mini-spray dryer.⁵

Implications of Recent Bench-Scale Work

In conjunction with the rapid development of in-duct SO₂ removal concepts, bench-scale evaluations have been conducted at the University of Texas and the Research Triangle Park (EPA-RTP) to investigate the fundamental chemistry and SO₂ capture mechanisms with hopes of ultimately improving the SO₂ removal and process economics of both spray drying and in-duct SO₂ removal concepts.

Recent work by Jozewicz and Rochelle⁶ has significant implications for improved dry SO₂ removal in that fly ash may play an important role in improving SO₂ removal. It is proposed that fly ash and calcium dissolve and react to form calcium silicate and calcium aluminum silicates which have very high (200-400 m²/g) surface areas. Further unpublished work using a sand-bed reactor by Jozewicz at the EPA-RTP indicates that very

high sorbent reactivities may be achieved using non-fly ash silica sources compared to commercial and reagent grade lime.

Other bench-scale efforts at EPA-RTP by Jorgensen, et al., also show improved sorbent reactivities using deliquescent additives such as calcium chloride and calcium nitrate, as well as the importance of duct conditions such as temperature and relative humidity⁷.

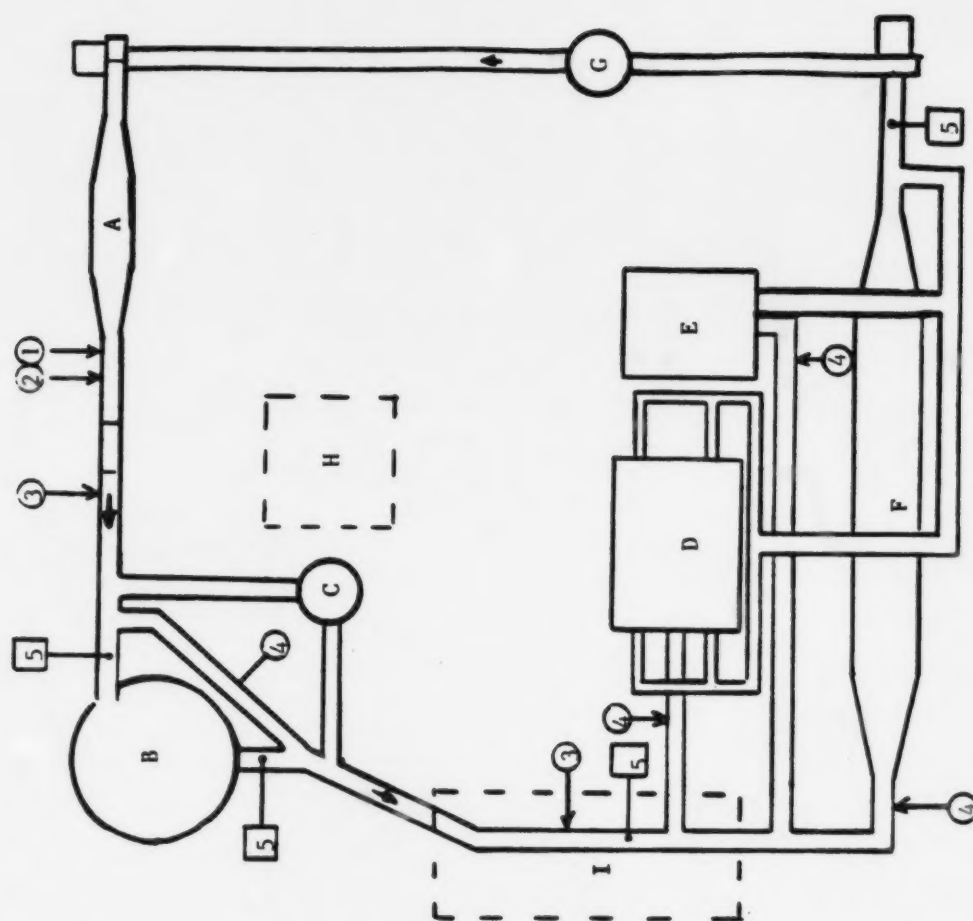
Laboratory work at Conoco also points out the need for improved sorbents and process concepts. Yoon, et al., report that, with sorbent reactivities similar to that of calcium hydroxide, slurry droplets would remove 10-20 percent SO₂ in the duct and another 20 percent SO₂ in an ESP of 400 ft²/1000 acfm (122 m²/1000 m³/min) collection area. Thus Conoco embarked on a program to develop a highly reactive sorbent for in-duct SO₂ removal and concluded that hydrated lime (surface area) and sodium carbonate (mass transfer additive) at a 9:1 molar ratio represented an effective sorbent. Field tests later showed up to 75 percent SO₂ removal using this sorbent in conjunction with Coolside humidification⁸.

The above developments indicate a need for further development work to (1) determine the most cost-effective sorbents with and without mass transfer additives, (2) establish the best process option which can take advantage of the enhancement of SO₂ removal by the reaction of calcium with fly ash (or another inexpensive source of silica), and (3) clearly identify the tradeoffs of injecting calcium as a dry powder followed by humidification/wetting versus injecting calcium as a slurry.

Proposed Research and Development at RTP

EPA has recently established a pilot dry flue gas cleaning facility with a supporting bench-scale laboratory to optimize both dry duct injection and spray drying processes in conjunction with conventional particulate matter controls as shown in Figure 1. Not shown in Figure 1 are the supporting laboratory reactors and a 50 cfm (1.42 m³/min) dry flue gas cleaning pilot plant used for sorbent screening and optimization work.

The laboratory scale reactors include two sandbed reactors described in Reference 6 which operate at low, 150-200°F (83-111°C), and high, 200-400°F (111-222°C), temperature, respectively, to allow comparison of reagent reactivities for longer (1 min-1 hr) sorbent residence times encountered in fabric filters. A differential flow reactor has been developed to allow comparison of sorbent particle reactivities for short (0.5-5 sec) residence times typically encountered in ductwork and ESPs. Currently, many calcium sorbents--including commercially available limes (slurried and pressure hydrated), lime/fly ash, lime/silica, lime/alumina, and lime + mass transfer additive combinations--are being evaluated on the laboratory reactors to determine the most promising sorbent and additive combinations to be evaluated on the larger scale pilot plants.



LEGEND:

- A - Air Heater
- B - Spray Dryer
- C - Mini-Spray Dryer/Humidifier
- D - Reverse Air Baghouse
- E - Pulse Jet Baghouse
- F - Electrostatic Precipitator
- G - Condenser/Scrubber
- H - Reagent Preparation Area
- I - Recycle Solids Area

- 1. Fly ash injection
- 2. Gas component injection
- 3. Steam injection
- 4. Sorbent injection
- 5. SO₂/O₂ CEM sampling point

Figure 1. EPA Dry FGC Facility, RTP, North Carolina.

The 2000 cfm (57 m³/min) pilot plant affords a unique opportunity to compare most of the previously discussed in-duct concepts with conventional spray drying over a wide range of process conditions and using either fabric filtration or electrostatic precipitation as the collecting medium. Best reagents identified from bench-scale efforts could then be tested according to the matrix presented in Table 1.

Figure 1 shows the current dry FGC pilot plant at RTP. All four process concepts shown in Table 1 can be investigated with the existing pilot plant. Parallel streams are available for: a conventional 2000 cfm (57 m³/min) Niro spray dryer, dry duct injection followed by water/steam humidification, dry injection with rewetting, and a recently added mini spray dryer with a 2 sec gas residence time. The mini spray dryer may also serve as a water-spray humidifier where dry sorbent is injected upstream and rewetted, or injected downstream in a humidified duct. Downstream solids collection may be via an electrostatic precipitator, reverse air baghouse, or pulse jet baghouse. All collected solids are conveyed to a central collecting area and are available for return to the reagent preparation area for recycle to the process.

The system uses a natural gas heater, and the desired fly ash and gaseous components are added downstream to simulate a wide variety of coal combustion flue gases. Initial gas conditions (including temperature and humidity) are well controlled because essentially 95 percent of the gas is passed through a final scrubber/condenser (where all remaining SO₂ and particulate matter are removed) and recycled to the gas burner. The only added air is that necessary to combust natural gas and raise gas temperatures to the desired level.

For retrofit work, lower cost, dry injection techniques coupled with ESP control will be emphasized. In addition to the factors affecting SO₂ capture, the downstream impact on the ESP should not be overlooked. One of the implicit goals of combined SO₂/particulate matter control is to maintain ESP performance. The precipitator in Figure 1 is currently undergoing a series of preliminary tests in order that it may be modeled and be amenable to cost-effective upgrading should future SO₂ testing indicate the need for improved performance. Manual samples upstream and downstream of the ESP will be taken during all test runs to ensure that particulate matter control is not compromised by SO₂ removal. Additional research on the impact of the increased grain loading due to SO₂ sorbent injection will be conducted in the 1,000 cfm (29 m³/min) ESP pilot plant also located at RTP. Concepts that will be evaluated in the ESP pilot plant include pulse energization, large diameter discharge electrodes, and multistage ESP.

Cost Advantages of Mini Spray Dryers

Compared to conventional spray dryer-FGD systems, the mini spray dryer concept offers a capital cost savings primarily in the absorber vessel. We compared costs for two retrofits: a 200 MWe utility boiler

TABLE 1. PILOT PLANT RESEARCH MATRIX FOR DRY FLUE GAS CLEANING

Process Concept:	Flue Gas Relative Humidity			External Ca/S Ratio			Sorbent/Additive Combinations				Inlet SO ₂ Conc., ppmv		
	40%	50%	60%	0.6	1.0	1.4	A	B	C	D	500	1000	2000
1. Conventional Spray Drying:													
a. with ESP													
i. once-through													
ii. reslurry solids & recycle													
iii. pressure hydrate solids + recycle													
b. with fabric filter (repeat ESP series)													
2. Mini Spray Drying (repeat conventional spray drying test sequence)													
3. Dry Injection With Rewetting (repeat spray drying test sequence)													
4. Dry Injection With Duct Humidification (repeat spray drying test sequence)													

and a 40 MWe-equivalent industrial boiler, both using 2 percent sulfur eastern coal. The differences between utility and industrial FGD operation affect both capital and operating costs: the major capital differences are the reagent inventory, sparing or redundancy in equipment, and waste solids transport and disposal. The most distinctive capital cost item is the absorber vessel. For the utility boiler, three vessels (one spare) were assumed, while no spares were assumed for the industrial boiler. For both boilers a 30 percent retrofit cost was assumed for relocation of existing ductwork and loss of power during construction.

Philosophically, the comparison of costs should not penalize a full spray drying system which has been engineered for high performance and low maintenance. For this reason all cases examined use the same process controls and auxiliary equipment as the spray drying system. Therefore the cost comparisons strictly reflect the difference in the absorber vessel and its operation.

For the conventional spray drying cases, spray dryer vessels of 10 sec residence time with rotary atomization are used, while the mini spray dryer consists of 3.3 sec residence time vessels (or duct expansions) with two-fluid nozzle atomization. Tables 2 and 3 illustrate the capital costs for these concepts applied to flue gas from a 2 percent sulfur eastern-coal-fired boiler. Costs for both once-through and product recycle are shown. Calculations were based upon the methodology developed for utility spray dryers by TVA⁹ and industrial spray dryer FGD systems by Radian Corporation¹⁰ and indexed to 1985 based on Chemical Engineering cost indices. Estimated retrofit costs are loosely based upon Reference 11, where a 30 percent retrofit capital charge was commonly observed.

TABLE 2. CAPITAL COST COMPARISONS FOR 200 MWe
UTILITY RETROFIT SO₂ CONTROL (1985 \$)*

	<u>Conventional Spray Drying</u>		<u>Mini Spray Drying</u>	
	<u>Once-Through</u>	<u>Recycle</u>	<u>Once-Through</u>	<u>Recycle</u>
Range of Total Capital Invest- ment x \$10 ⁶	29.4 - 41.2	35.4 - 47.2	18.1 - 29.9	24.6 - 36.4
\$/kW	147 - 206	177 - 236	91 - 150	123 - 182

TABLE 3. CAPITAL COST COMPARISONS FOR 40 MWe
INDUSTRIAL RETROFIT SO₂ CONTROL

	<u>Conventional Spray Drying</u>		<u>Mini Spray Drying</u>	
	<u>Once-Through</u>	<u>Recycle</u>	<u>Once-Through</u>	<u>Recycle</u>
Total Capital Investment x \$10 ⁶	4.18	5.28	2.44	3.74
\$/kW	105	132	61	94

A retrofit of conventional spray drying to a 200 MWe utility boiler would range from \$147 to 236/kW depending on the extent of reworking the existing facility and whether recycle of dryer solids is practiced. A mini spray dryer at the same facility would range from \$91 to 182/kW but in every case would represent a savings of about \$11 million in retrofit capital costs.

Similar savings are shown for a 40 MWe equivalent industrial boiler retrofit, though the actual costs per kilowatt are somewhat lower than for the utility. This apparent violation of the economies of scale is due to the industrial application's having no spare modules and several other aspects such as reagent handling and waste solids handling and disposal which tend to lower capital expenditures and increase operating costs. Moreover the industrial project incurs less interest on capital due to an 18 month construction period, compared to 36 months for the utility retrofit.

Operating costs are not presented; the information needed to develop these costs will not be available until the mini dryer concept has been evaluated. Operating costs will necessarily be sensitive to the type and quantity of reagent used. Similarly cost-effectiveness will be highly dependent upon the sorbents chosen and whether product recycle is used.

For conventional spray drying retrofit we feel that, for low to medium sulfur coals (up to 3% S), 70 percent SO₂ removal could be expected for a once-through system and 90 percent SO₂ removal for fly ash recycle [at reasonable (1.0-1.4) calcium-to-sulfur ratios and saturation approach temperatures of 18-27°F (10-15°C)] based upon previous experience^{12,13}. Goals of the mini spray drying concept should be 50 percent SO₂ removal for once-through and 70 percent SO₂ removal for product recycle. Therefore, while the capital cost advantages of mini spray drying are apparent, the operating costs and relative cost-effectiveness advantages will depend largely upon sorbent economics, remaining boiler life, and the size of the boiler.

Impact of Mini Spray Dryer on Particulate Emissions

The EPA/Southern Research Institute ESP computer model was used to estimate the impact of a mini spray dryer collecting SO₂ from a plant burning 2% S eastern bituminous coal. ESP performance was estimated for Ca/S ratios ranging from 0.6 to 1.4. The specific collector area (SCA) of the base ESP was assumed to be 210 ft²/1000 acfm (64 m²/1000 m³/min) and was capable of meeting an emission standard of 0.1 lb/10⁶ Btu (43 ng/J). This emission limit is representative of the particulate emission standard that most older plants have to meet. The SCA for the mini spray dryer cases was 245 ft²/1000 acfm (75 m²/1000 m³/min). The SCA for the mini spray dryer cases increased because the flue gas temperature decreased from 300°F (167°C) for the baseline to 160°F (89°C) for the mini spray dryer cases.

Gas cooling and humidification from the mini spray dryer will also reduce the electrical resistivity of the fly ash/sorbent mixture to a value suitable for collection in an ESP. Limited laboratory resistivity measures taken for this analysis show that the electrical resistivity of reacted lime/fly ash mixtures drops from about 2x10¹² ohm-cm at 300-320°F (167-178°C) to less than 2x10¹⁰ ohm-cm at 160°F (89°C). This low value of electrical resistivity will provide good electrical conditions in the ESP.

The ESP model results showed that the particulate matter emissions for a 2% S coal and Ca/S ratios less than 1.5 would meet the particulate emission standard of 0.1 lb/10⁶ Btu (43 ng/J). The calculations also showed that the plume opacity would be under 20 percent in all cases. Particulate emissions for higher Ca/S ratios and higher sulfur coals would exceed the 0.1 lb/10⁶ Btu (43 ng/J) limit and some particulate retrofit work would be required.

The multistage ESP retrofit discussed by Sparks et al.⁴ would ensure that the particulate emissions from a mini spray dryer system did not exceed particulate emission limits. The multistage ESP retrofit might add \$5-10/kW to the capital costs of a retrofit mini spray dryer system.

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ELECTROSTATIC PRECIPITATOR PERFORMANCE CHARACTERIZATION
IN SPRAY DRYER FLUE GAS DESULFURIZATION SYSTEMS

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ABSTRACT

With the advent of potential acid rain legislation on a Federal level, along with the new EPA rulings on tall stacks for pollutant dispersion, many utilities are beginning to evaluate possible emissions reduction options for existing power plants. One option that may be cost effective is the installation of a spray dryer absorber (SDA) ahead of a plant's existing electrostatic precipitator (ESP). In order to properly assess the impact of SDA placement ahead of an existing ESP, a data base is required which characterizes the effects due to variations in ESP design and SDA operation on particulate removal. In order to accumulate such a data base, Tennessee Valley Authority (TVA) and Combustion Engineering (CE) have initiated a program to study the impact of key design factors and parameters.

The program is being carried out at TVA's Shawnee Test Facility using CE's mobile SDA and ESP pilot plants. Testing is focused on collecting ESP and SDA operating/performance data for a variety of ESP hardware configuration. The data will be analyzed by CE in two different ways: (1) with the Matts-Ohnfeldt modified Deutsch equation, and (2) the Southern Research computerized model. The effects of varying discharge electrode design, plate spacing, SCA, and SDA process conditions are under investigation.

The program was initiated in August, 1985, and will conclude in June, 1986. A program plan, equipment description, and preliminary results are described.

INTRODUCTION

RATIONALE FOR CURRENT STUDY

With the advent of potential acid rain legislation on a Federal level, along with revised EPA rulings on credits previously allowed for tall stacks, some utilities are beginning to re-evaluate possible emissions reductions options. In the past, the major options considered viable by utilities burning medium to high sulfur coal in the affected areas were coal switching, load switching, and wet scrubbing. However, in a recent study (1), it was shown that the installation of one or more spray dryer absorbers (SDA's) ahead of a plant's existing electrostatic precipitator (ESP) may be a cost effective measure for plants burning medium to high sulfur coal east of the Mississippi River.

To properly evaluate the retrofit option for placing an SDA ahead of an existing ESP and still achieving the particulate removal requirements, the Tennessee Valley Authority (TVA) and Combustion Engineering (CE) have joined resources to collect the data necessary to formulate a sound data base.

STUDY OBJECTIVES

The test program is designed to accumulate and analyze data on ESP performance after a hopperless SDA. The SDA vessel is hopperless in so much as there is no ash classification within the vessel nor is any ash removal equipment required. All the ash and solid reaction products are removed in the ESP hoppers. Parameters such as electrode design, plate spacing, SCA, and process operating philosophy are being evaluated. Two different ESP performance models are being used for data interpretation: the Deutsch equation with the Matts-Ohnfeldt modification (2), and the Southern Research Institute (SoRI) computer code (3). The objective is to provide guidelines for predicting efficiencies of full-scale equipment.

DESCRIPTION OF ANALYTICAL APPROACH

A major objective of the data analysis is to project full scale ESP performance using two models: the modified Deutsch equation and the SoRI computer code. The modified Deutsch equation requires particulate collection efficiency, gas flow, and plate area data.

$$1-\text{efficiency} = \exp[-SCA \times W_e \times .001968]^k$$

where

SCA = specific collection area in square feet plate/1,000 acfm

W_e = effective modified migration velocity, cm/sec

.001968 = units conversion factor

k = empirical exponent

As can be seen from the above equation, there are two coefficients, W_e and k , which must be derived from the experimental data. The data being collected in this program will be used to study the variation in W_e and k as plate spacing, electrode design, SCA, and process operating conditions are varied. Absolute values of W_e and k derived from this pilot unit are not expected to be directly relatable to those for a full scale ESP. However the relative changes in W_e and k between operation with and without the SDA should be quite useful for predicting W_e and k for full scale units with retrofit SDA's.

The SoRI computer model projects the efficiency of an ESP given certain inlet conditions and ESP configuration. The input data for any fixed set of inlet conditions is listed below:

Input	Units	Source
Inlet mass loading	g/acf	Test data
Particulate density	kg/m	Constant = 1
Dielectric ratio	—	Set as constant — has negligible effect in model
Ion mobility	m ² /V/s	Empirical, based on gas composition
Resistivity or V/I data	ohm-cm	Test data
Particle size distribution	μm (mmid)	Test data
Size distribution bands	%	Test data
Gas viscosity	kg/m/s	Test data
Plate length	ft	Hardware data
Plate area	ft ²	Hardware data
Total wire length	ft	Hardware data
Corona wire radius	in	Hardware data
Number of wires	#	Hardware data
Wire-to-plate spacing	in	Hardware data
Wire-to-wire spacing	in	Hardware data
Gas volume flow rate	acfm	Test data
Gas velocity	ft/s	Test data
Gas temperature	deg F	Test data
Gas pressure	atm	Test data

Characteristics of the pilot ESP are inputted and the ion mobility adjusted by repetitive execution until the predicted efficiency matches the measured efficiency. A range of ion mobilities may be determined for the process test conditions. Particular attention is being focused on the effect of relative humidity. However, apparent ion mobility may also be affected by variations in hardware configurations or SCA.

In summary, then, the SoRI computer model will be calibrated with the ESP pilot data and subsequently used to predict the performance of full-scale units currently under scrutiny. Changes in ion mobility will be useful in predicting the effect of full scale retrofit SDA's on the operation of existing ESP's.

TEST EQUIPMENT DESCRIPTION

ELECTROSTATIC PRECIPITATOR

The pilot ESP is supplied by Combustion Engineering (CE) and consists of three fields. The collecting plates are five feet high, and the total treatment length is 14.4 feet. The discharge electrodes are of CE's framed design and are capable of being outfitted with either round wires, barb

wires, star wires, or rigid electrodes. Duct spacing is modifiable to either 9, 12 or 16 inches. The collecting plates and wire frames are rapped from the top using pneumatic hammers.

Auxiliary equipment for the ESP includes the following: three (3) 75 KV transformer/rectifier sets, three CE Model 80 automatic voltage controllers complete with control panel, a flexible screw conveyor providing ash removal for each of the three hoppers serving the three fields, and rotary airlocks on each hopper.

Figure 1 is a photograph of ESP as installed at the test site in Paducah, Kentucky. The ESP is mounted on a 362-1/2 by 108 inch flatbed trailer which readily facilitates transport and installation. The inlet and outlet ductwork is arranged to allow test crews space for collecting mass and impactor train samples.

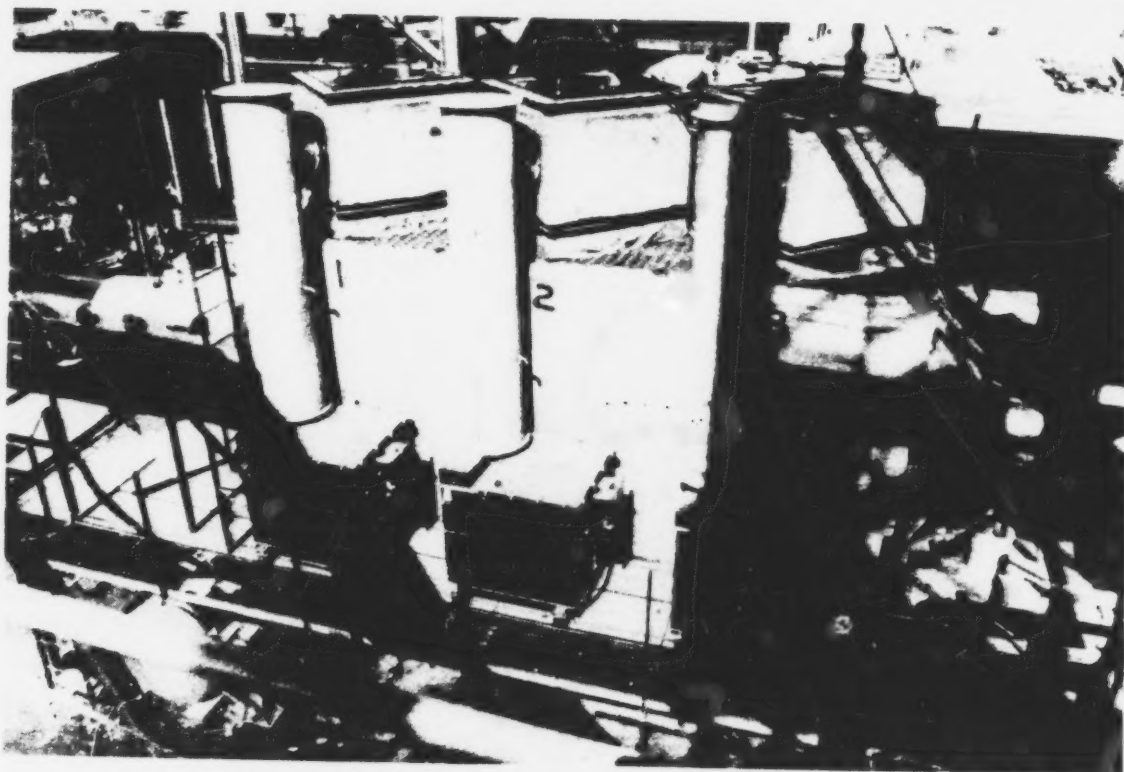


Figure 1. Photo of ESP at Shawnee Steam Plant

Reproduced from
best available copy.



This ESP facility was previously located at the Gadsden Steam Plant of Alabama Power Company, where it was employed in a detailed test program to evaluate electrode configurations and duct spacing effects.

SPRAY DRYER ABSORBER

The SDA is also supplied by CE. It is a standard Bowen Engineering design (1979) constructed of unlined and uncoated carbon steel. The inside diameter of the SDA is 7 feet at the widest point, with a height of the cylindrical section being 46 inches. Below the cylindrical section, the SDA is equipped with a 60 degree cone (measured from the horizontal) and has a total

internal volume of 224 cubic feet. Flue gas enters the SDA through a swirler/diffuser at the top and exists at the bottom of the conical section. All the dried droplets from the rotary atomizer are entrained with the incoming flyash in the flue gas and pass on to the ESP.

Atomization of the slurry in the SDA is accomplished with a Bowen Model AA6, 10-HP spray machine. The spray machine is powered by a solid-state frequency inverter providing an atomizer speed range from 0 to 22,500 RPM. The atomizer wheel is an abrasion resistant, nozzle type. It is 6 inches in diameter and is fitted with three 0.25 inch ID silicon carbide orifices.

Figure 2 is a photograph of the SDA along with some of the other auxiliary equipment required for the test program. Together with some of the additive preparation equipment and the instrumentation and controls, the SDA is mounted on a 40 foot long split-bed trailer, which, as with the ESP, readily facilitates transportation and setup at field sites.

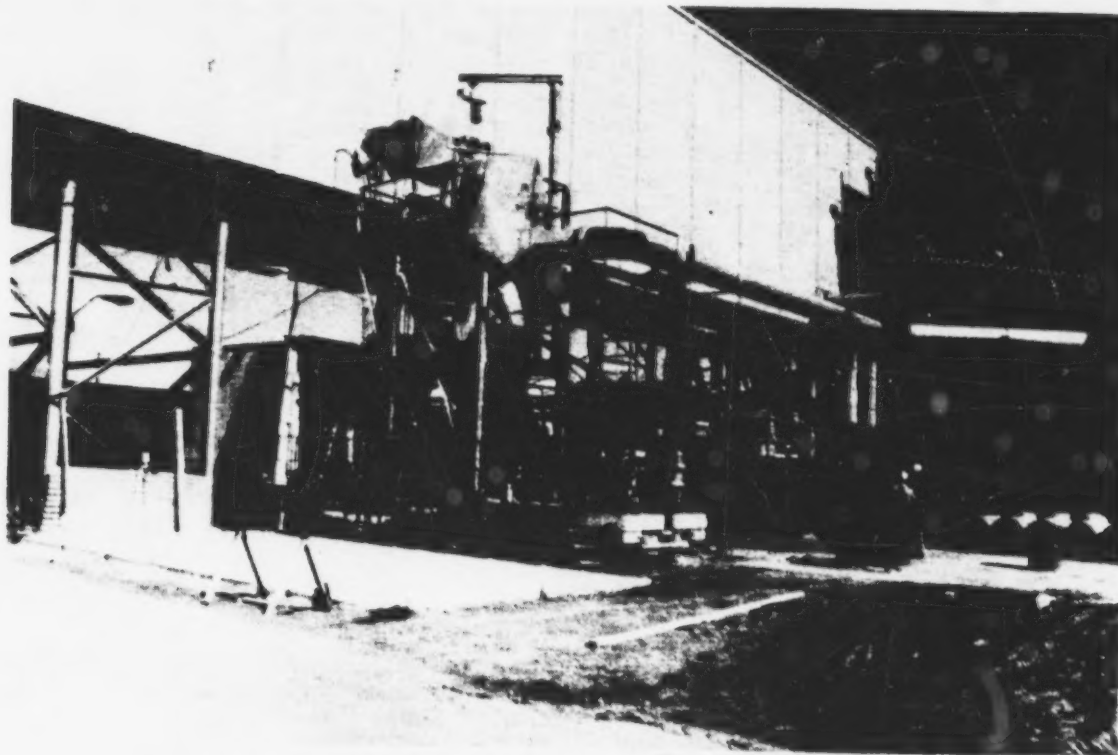


Figure 2. Photo of SDA at Shawnee Steam Plant

LIME SLURRY PREPARATION AND HANDLING

Figure 3 is a schematic of the stream flow with the major equipment. Pebble quicklime is loaded from 50-lb bags intermittently as required during the slaking periods. The lime is slaked in a Wallace and Tiernan Series A-758, 1,000 lb/hr paste slaker. Slaked lime slurry of 25% solids is accumulated in a 1,000 gallon tank and transferred continuously from this tank to a 400 gallon additive feed tank. Lime slurry is pumped from this tank and is mixed in-line with recycle slurry at a rate sufficient to meet the stoichiometric ratio requirements of the test program. The mixed slurry passes on to the spray machine.

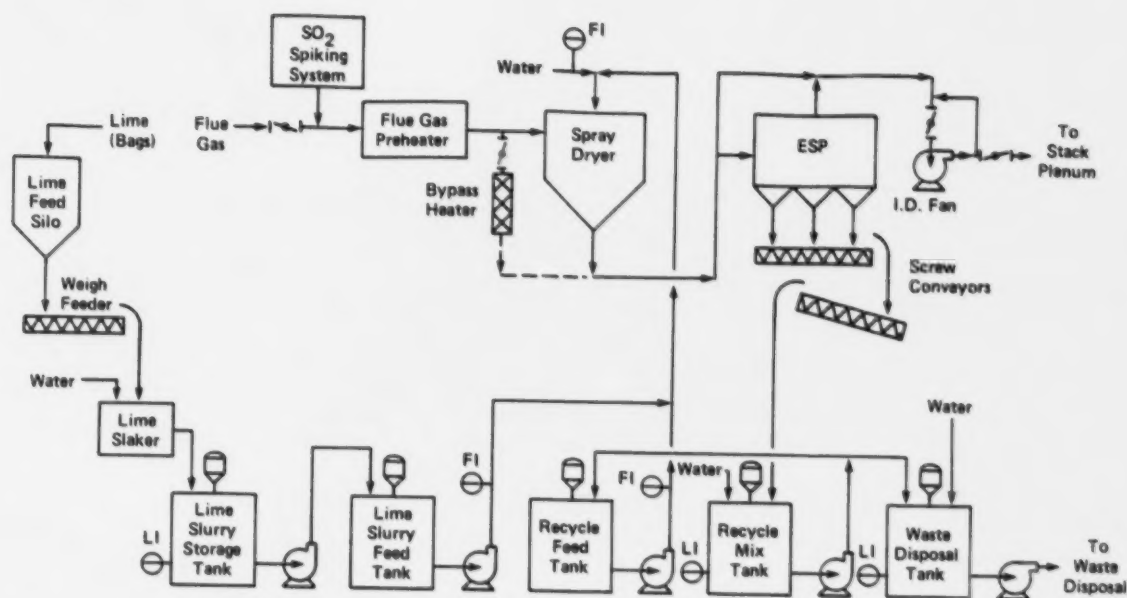


Figure 3. Schematic of Pilot Plant Process Flow and Equipment

The recycle slurry preparation system provides a means for removing the excess flyash and reactant products from the system. Waste solids from the ESP are conveyed to a 550 gallon tank where they are slurried to 35% solids. From here, recycle slurry is pumped either to a 560 gallon recycle slurry feed tank or a 470 gallon waste slurry disposal tank. Slurry in the waste disposal tank is further diluted as required to keep the velocity above three feet per second in the waste slurry discharge line. Slurry from the recycle feed tank is mixed in-line with the lime slurry in an amount sufficient to maintain a desired SDA outlet gas temperature.

GAS PATH AND SO₂ SPIKING

Flue gas is obtained from Unit 10 of TVA's Shawnee Steam Plant. The inlet duct runs from a point ahead of Unit 10's existing fabric filter, goes along outside the fabric filter, then runs to the inlet of the SDA. Sulfur dioxide liquid is injected and vaporized in the duct ahead of the SDA as required to maintain an inlet concentration of 2,800 PPMV. Ahead of the injection point, SO₂ concentration varies from about 2,000 to 2,600 PPMV. SO₂ concentration is measured via an auto-switching sampling system and a DuPont Model 460 UV Analyzer. Sampling points are located before the SDA, after the SDA but ahead of the bypass injection point, ahead of the ESP and after the ESP.

The flue gas from the duct is routed through an electrical resistance heater to maintain a constant SDA inlet temperature of 300 degrees Fahrenheit. A bypass line around the SDA is also outfitted with an electrical resistance heater to replace heat lost via conduction when gas bypassing is employed.

Flue gas is drawn through the pilot plant system by a 75-HP Spencer Turbo-Blower located in the gas path after the ESP. Discharge flue gas is routed to the steam plant's stack plenum.

TEST PROGRAM

HARDWARE CONFIGURATIONS

Five ESP hardware configurations were chosen which were felt to best represent the types of configurations presently installed in commercial systems of interest or which had good potential for ESP retrofit application. The standard configuration is round wire discharge electrodes with a plate spacing of nine inches, which was chosen as the first configuration. The other configurations of interest were at wide plate spacings. A smaller SCA will result when plate spacing is changed to 12 or 16 inches and the number of gas passages are reduced. Different discharge electrodes would be required to help facilitate particulate collection. Two different discharge electrodes (star and barb) were chosen for each of the two wider plate spacings. In summary, the configuration are as follows:

Configuration	Plate Spacing, inches	Discharge Electrode
1	9	0.1015 in. Round Wires
2	12	Star Wires
3	12	Barb Wires
4	16	Barb Wires
5	16	Star Wires

An additional advanced concept that was considered was intermittent energization. However, the range of ash resistivities anticipated did not warrant very strong consideration, and plans for evaluating this concept were dropped not long after testing began.

PROCESS CONDITIONS

The process conditions were chosen to evaluate variations in SCA and ash characteristics, specifically those ash characteristics which vary as a result of practical SDA variances in operation. Correspondingly, then two sets of gas flows and SDA operating conditions have been chosen to form a test matrix, as follows:

Condition	Target ACFM at	
	Outlet of ESP	SDA Operating Condition
1	3,000	SS-1
2	3,000	SS-2
3	3,000	Off
4	4,000	SS-2
5	4,000	SS-1

The SDA operating conditions SS-1 and SS-2 are further described below:

SDA Operating Condition	SS-1	SS-2
SDA Inlet SO ₂ , ppmv	2,800	2,800
SDA Inlet Temp., °F	20	30
Stoichiometric Ratio, Mol/Mol	1.2	1.0

Predicted material balances were prepared for each process condition. These are provided on Table 1.

TABLE 1. PREDICTED MATERIAL BALANCES FOR PHASE I

SDA Operating Condition	SS-1	SS-1	SS-2	SS-2
ACFM at ESP outlet	3,000	4,000	3,000	4,000
EXPECTED FLOWS, lb/hr				
Flue gas to SDA	11,400	15,200	11,400	15,200
Flue gas to ESP	11,800	15,700	11,800	15,700
Flue gas from ESP	12,100	16,000	12,100	16,000
Solids from ESP	303	416	296	396
Recycle mix slurry	866	1,188	846	1,131
Recycle mix to feed	180	240	219	292
Waste slurry, 10% solids	2,480	3,320	2,190	2,940
Lime slurry	405	540	338	450
EXPECTED FLOWS, GPM				
Recycle transfer rate	1.39	1.93	1.35	1.82
Waste disposal rate	4.63	6.2	4.09	5.49
Recycle feed rate	0.29	0.39	0.35	0.47
Lime feed rate	0.7	0.79	0.58	0.78
TEMPERATURES, F				
SDA inlet	300	300	300	300
SDA outlet	147	147	156	156
Saturation (SDA outlet)	127	127	126	126
ESP outlet	131	132	130	130
SPECIFIC GRAVITY				
Lime slurry	1.158	1.158	1.158	1.158
Recycle slurry	1.236	1.236	1.236	1.236
Waste slurry	1.070	1.070	1.070	1.070

SCHEDULE

Around the clock operation was specified initially to allow time for the ESP ash layer to condition thoroughly. Three days of operation at the desired process condition were provided before ESP performance data was collected. Presently, however, this has been reduced to two days to conserve operating resources. Testing began in early August, 1985, with tests of seven day duration and shortened to five day duration in December.

The order of testing is summarized below:

Test ID	Hardware Configuration	Process Condition
CE1-1-01	1	1
CE1-1-02	1	3
CE1-1-03	1	2
CE1-1-04	1	4
CE1-1-05	1	5
CE1-2-01	2	1
CE1-2-02	2	3
CE1-2-03	2	2
CE1-3-01	3	1
CE1-3-02	3	3
CE1-3-03	3	2
CE1-4-01	4	1
CE1-4-02	4	3
CE1-4-03	4	2
CE1-5-01	5	1
CE1-5-02	5	3
CE1-5-03	5	2

DATA COLLECTION AND PROCESSING

Process data is recorded once per hour by the system operators, and this data is transcribed to floppy diskette once per day. The process data collected is listed on Table 2. Voltage-current curves (V-I) are recorded once per day by site engineers. An on-site analytical laboratory provides most analytical support. Table 3 summarizes the sampling and analyses performed.

Important test data needed for the SoRI model comes, in part, from mass train, impactor, and in-situ resistivity testing conducted on a regular schedule by a TVA test team. Mass train tests are conducted generally on days 2 and 3 of testing, Impactors on days 2 and 4, and in-situ resistivities are taken on days 3 and 4.

PRELIMINARY RESULTS

ASH CONVEYOR

The flexible screw conveyor taking flyash from the ESP hoppers, while installed according to the manufacturer's specification, has not performed very reliably and has accounted for some lost test time. The major problem with this unit has been binding between the flexible screw and the channel wall. Binding has resulted in numerous incidents of unwinding of the screw as well as overheating and burnout of the drive motor. A portion of this conveyor has been replaced with a new, separate conveying system to correct this problem.

SPRAY MACHINE AND AUXILIARIES

The spray machine experienced several problems, especially with the rather concentrated slurry initially fed to the slurry distributor. An early test plan called for increased recycle solids rates by running more concentrated recycle slurries than the currently specified 35%. Running recycle slurry concentrations from 45 to 55% caused buildup in the spray machine slurry distributor and on the atomizer wheel itself. Slurry build-up caused atomizer imbalance and some rubbing, resulting in the premature failure of spray machine bearings. In addition, slurrying and

handling tests high solids concentrations posed problems with the current equipment at the facility.

Some reliability problems were experienced with the Teledyne frequency inverter, a unit which has historically been quite reliable. Due to its complexity, long lead time for service, and lack of spare parts, failures in the frequency inverter have caused longer down times than normally expected for equipment of this type.

TABLE 2. LIST IDENTIFYING
THE PROCESS DATA COLLECTED

TEMPERATURES, F

Preheater inlet	SDA inlet
SDA outlet (dry bulb)	SDA outlet (wet bulb)
ESP inlet	ESP outlet (dry bulb)
ESP outlet (wet bulb)	SDA bypass
ESP bypass	Atomizer slurry feed
Ambient	

PRESSURES, in. H₂O

Preheater inlet	SDA inlet
SDA outlet	SDA differential
ESP outlet	ESP differential
System differential	ID fan outlet
ID fan differential	Barometric (in. Hg)

PRESSURES, psig

Lime slurry feed pump	Recycle feed pump
Recycle mixer slurry pump	Waste disposal pump
Lime slurry transfer pump	SO ₂ cylinder

FLOW RATES

ESP outlet flue gas, acfm	Recycle slurry feed, GPM
Lime slurry feed, GPM	Trim water, GPM

GAS CONCENTRATION (ppmv SO₂ and O₂)

SDA inlet	SDA outlet
ESP inlet	ESP outlet

MISCELLANEOUS

Opacity at ESP outlet	Tank levels (all tanks)
Atomizer speed	Boiler load

ESP ELECTRICAL (all 3 fields)

Primary voltage	Primary current
Secondary voltage	Secondary current
Estimated spark rate	

TABLE 3. TABLE OF SAMPLES AND ANALYSES PERFORMED

Material	Dry Lime	Lime Slurry	Recycle Slurry	ESP Solid Product	coal	Unit #10 Flyash
Reactivity	a	---	---	---	---	---
Available						
Alkalinity	a	b,c	c	f	---	---
Ca ⁺⁺	a	b,c	f	f	---	---
CO ₃ ⁺	a	b,c	f	---	---	---
% Solids	---	b,c	c	---	---	---
Acid Insolubles	a	---	---	f	---	---
Loss On						
Ignition	a	---	---	---	---	---
Viscosity	---	c	c	---	---	---
pH	---	---	c	---	---	---
Particle Size						
Distribution	---	---	c	f	---	---
SO ₃ ⁺	---	---	f	f	---	---
SO ₄ ⁺	---	---	f	---	---	---
Mg ⁺⁺	a	c	---	f	---	---
Total Sulfur	---	---	---	f	---	---
% Moisture	---	---	---	f	---	---
E.P. Toxicity	---	---	---	g	---	g
Proximate	---	---	---	---	f	---
Ultimate	---	---	---	---	g	---

Legend: a. Once/delivery
b. On request or as required
c. Once/batch
d. Once/hour
e. Once/4 hours
f. Once/day
g. Once/week

PRELIMINARY CONCLUSIONS

While not all the program objectives have been achieved on the desired schedule, much quality data has been collected, and 6 test conditions have been completed by the end of December. With continued, diligent effort by test engineers and operators, program objectives should be achieved before the end of the planned test period. The data, results, and analyses from this study, when published, should be a significant contribution to public domain information on SDA retrofits.

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FABRIC FILTER OPERATING EXPERIENCE ON A DRY FGD SYSTEM
AT AUSTELL BOX BOARD CORPORATION

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ABSTRACT

AUSTELL BOX BOARD CORPORATION, a manufacturer of paper board from recycled wastepaper, installed a new pulverized coal fired boiler and a dry FGD system in late 1982. The boiler is rated for 250,000 lb/hr of steam at 250 PSIG and 406°F (saturated). The plant fires bituminous coal with an average sulfur content of 2.2%; an average ash content of 14%; and an average heating value of 12,000 Btu per pound. Emission requirement limits are 1.2 lb SO₂, 0.7 lb NO_x and 0.1 lb of particulate per million Btu. The FGD system has been in service since March of 1983. The dry FGD system consists of a spray dryer followed by a fabric filter. The fabric filter is a WHEELABRATOR-FRYE pulse jet unit utilizing on-line cleaning with a design air to cloth ratio of 4. Significant experience with various fabrics has been gained during operation of the unit. The fabric filter has operated with both woven and felted fiberglass bags. Tests began in February, 1985 utilizing full compartments of 4 different type felted bags. Results of these tests are presented in the paper. Overall operating experience with the fabric filter as well as with the pneumatic ash conveying system serving it are also presented.

INTRODUCTION

Austell Box Board Corporation, a subsidiary of Caraustar Industries, manufactures paper board for use in making paper tubes and folding/setup boxes. Sweetwater Paper Board Corporation, also a Caraustar subsidiary, manufactures paper board for the building industry. Together these companies operate three paper mills at the Austell GA location, all using 100% waste paper as raw material.

In December of 1982, a Babcock & Wilcox (B&W) 113,500 kg/hr (250,000 lb/hr) pulverized coal (with natural gas and #2 oil backup) fired boiler was brought on stream, replacing existing gas and oil fired package boilers.

The "back-end" equipment on the boiler consists of a spray dryer (utilizing quicklime for SO₂ capture) followed by a fabric filter. Both the spray dryer and fabric filter were supplied by Wheelabrator-Frye.

The boiler was fired with pulverized coal in mid-January, 1983. After obtaining stabilized coal firing conditions, the fabric filter was placed in service in early February, 1983. The spray dryer was placed in continuous service in mid-March, 1983.

EPA compliance testing performed in April of 1983 indicated that the system was well within the allowable SO₂ and particulate emission limits. The unit has passed two additional annual EPA compliance tests since the initial test.

EMISSION REQUIREMENTS

Emission regulations applicable at the time of Air Quality permitting required that we limit oxides of nitrogen (as NO₂) to 0.301 g/MJ (0.7 lbs/MBtu), sulfur dioxide to 0.517 g/MJ (1.2 lbs/MBtu), and particulate emissions to 0.0426 g/MJ (0.1 lbs/MBtu). Maximum opacity limit is 20% except for one 6 minute period each hour which may equal 27%. All limits except opacity are based on a 3 hour rolling average. Opacity limits are based on a 6 minute rolling average. Hydrocarbons and CO are required to be maintained at their lowest practicable minimums.

Continuous stack monitoring is required for SO₂, NO_x, and opacity. In addition, continuous O₂ or CO monitoring is required at the boiler outlet.

FUEL SPECIFICATIONS

The boiler is designed to fire pulverized coal, #2 fuel oil, and natural gas or any combination of these three. The primary fuel, of course, is coal with the other fuels being used for coal firing equipment maintenance, emergencies, and boiler start up. Design coal specifications

are shown in Table 1.

TABLE 1. DESIGN COAL SPECIFICATIONS

	As Received
Heating Value	
kJ/kg	26,723
Btu/lb	11,500
Moisture	10%
Volatile Matter	38%
Fixed Carbon	39%
Ash	13%
Sulfur	2.5%
Ash Softening	1204°C (2200°F)

Coal actually fired in the unit through November of 1985 has in general been of higher heat value ranging from 27,376 kJ/kg (11,800 Btu/lb) to 29,696 kJ/kg (12,800 Btu/lb) with an average sulfur of 2.2% and an average ash of 13%.

In December 1985, the coal specifications were changed to 12,500 Btu/lb, 1.0% average sulfur, and 10% average ash.

SYSTEM DESCRIPTION

The boiler is a B&W Stirling SPB design utilizing six burners and three pulverizers (two burners per pulverizer). The furnace is of balanced draft design utilizing a single forced draft fan and a single induced draft fan. Each pulverizer is capable of 40% of full boiler load. Normal operating pressure is 1722 kPa (250 PSIG) saturated. Flue gas heat recovery equipment consists of a single Ljungstrom regenerative air heater. Flue gas exit temperature (spray dryer inlet temperature) is controlled by a steam coil air heater on the makeup air side of the regenerative air heater.

FLUE GAS DESULFURIZATION SYSTEM

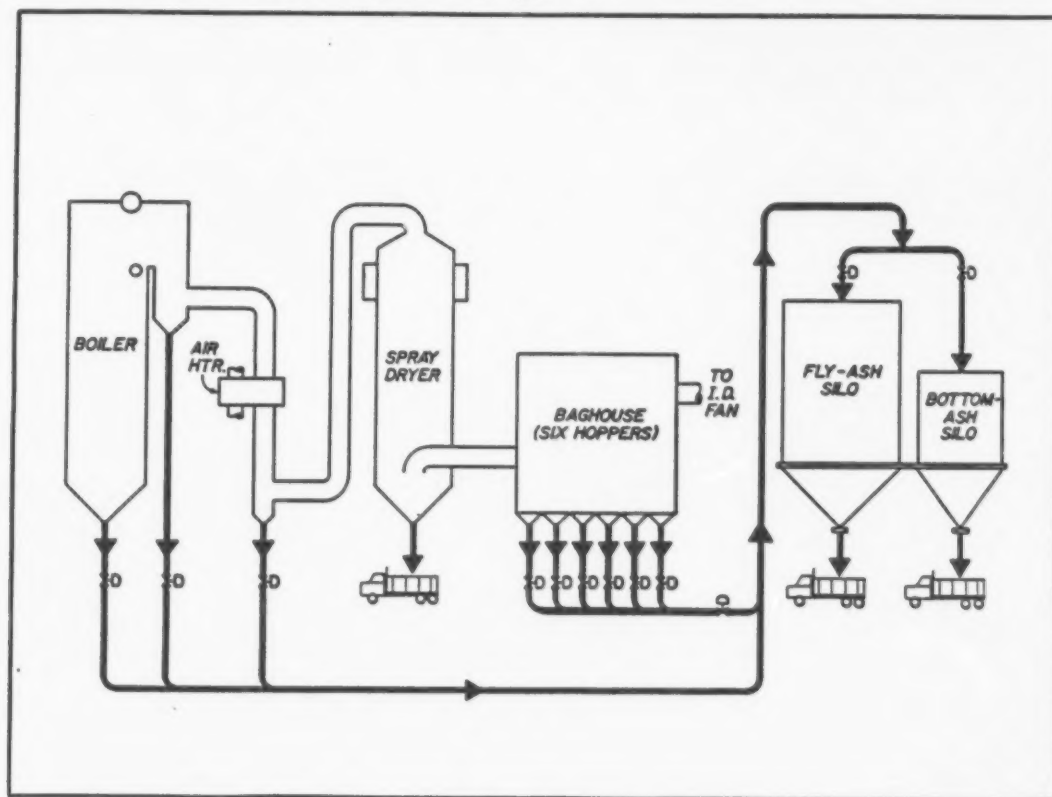
The flue gas desulfurization system (FGD) consists of a two fluid nozzle spray dryer, a pulse-jet fabric filter, and pebble lime storage, slaking, and feed systems. The spray dryer is a parallel flow design utilizing co-current downward gas flow. Flue gas volume at 100% boiler load is approximately 3113 m³/min. (110,000 ACFM at 176°C [350°F]). The drying chamber is 5.49m (18 feet) in diameter and provides a nominal 10 second residence time at full load. Flue gas and entrained particles

exit from the side of the dryer into the fabric filter. Large particles are discharged from the dryer bottom hopper.

The fabric filter is a 6 compartment pulse-jet unit. A total of 1530 152.5mm (6 inches) diameter by 3.05 meters (10 feet) long bags provide over 2230 square meters (24,000 sq ft) of filter surface area. Bag cleaning can either be controlled by pressure drop across the unit or by preselected time intervals. Gas to cloth ratio with all 6 modules on-line at full load is approximately 4 to 1. The fabric filter is furnished with 100% bypass capabilities which can be activated manually or will be activated automatically in the event of unnecessarily low or high temperatures entering the fabric filter.

ASH CONVEYOR

Fly ash is removed from the fabric filter hoppers by pneumatic conveyor and stored in a silo for transport to an on site landfill. Bottom ash from the boiler is conveyed by the same pneumatic conveying system to a separate silo. Ash from the spray dryer bottom discharge is currently being loaded into a special truck and hauled directly to a landfill. The total system ash flow is shown in Figure 1.



CFB FILE NO. A 100-030-1

Figure 1. Ash flow

PROCESS CONTROL

Both the boiler and spray dryer are being controlled by a distributed digital system supplied by Leeds & Northrup. Data gathering and report generation is being done by a DEC computer. DEC software and hardware were furnished by Automated Technologies.

The primary sensors for the liquid flow control system are the spray dryer temperature sensor and the stack SO₂ monitor. Total liquid flow to the spray dryer is controlled based on the dryer temperature. The percent slurry versus the percent dilution water is controlled through a ratio control scheme based on the stack SO₂ measurement as indicated in Figure 2.

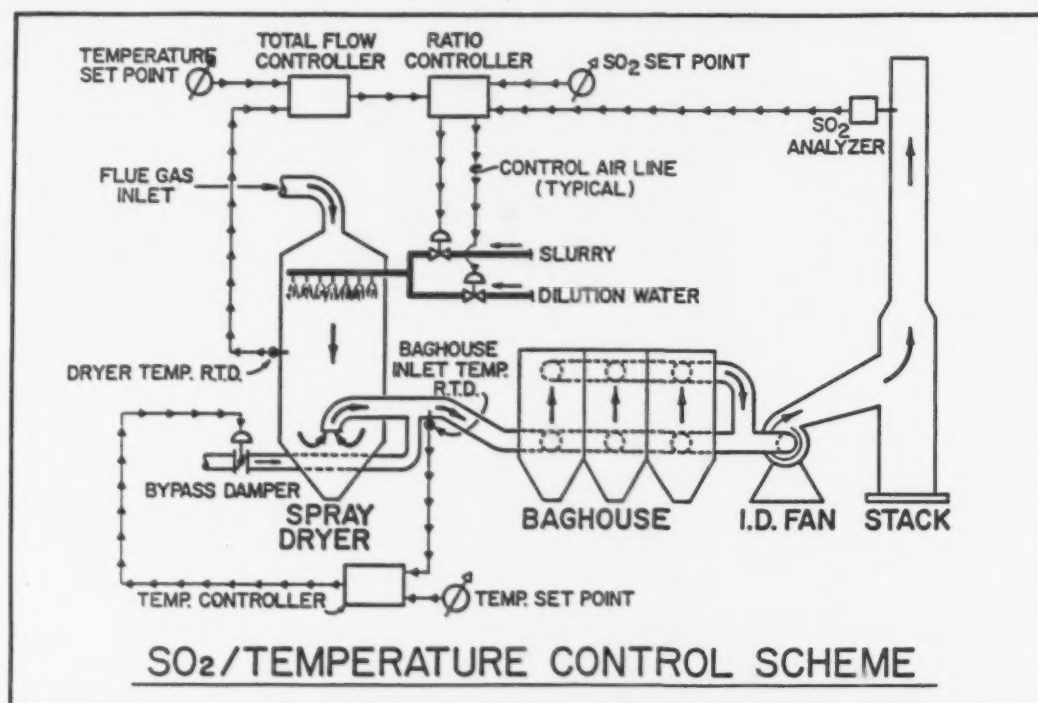


Figure 2.

Inlet temperature to the fabric filter is controlled by an independent loop. The inlet temperature signal is fed to a controller which positions the reheat damper to maintain the set point fabric filter inlet temperature. Maximum hot gas bypass is approximately 10% of the total gas flow. Design operating temperatures are shown in Figure 3. The spray dryer is currently being operated closer to 165°F to provide larger safety margins at the expense of poorer lime utilization.

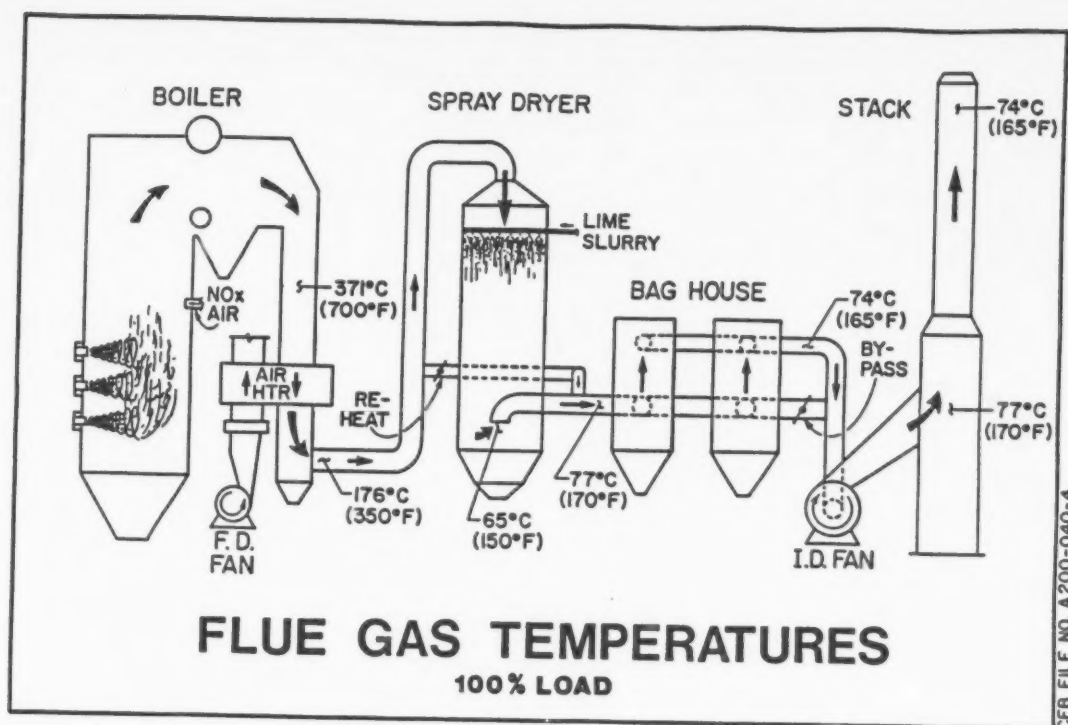


Figure 3.

OPERATING EXPERIENCE

FABRIC FILTER

The fabric filter was initially started up with felted fiberglass bags. The unit performed very well through December of 1983, with few bag failures. During this initial phase of operation, the spray dryer was going through its "shake-down" phase which resulted in several moisture upset conditions in the spray dryer and probable upsets in the fabric filter.

By mid-January 1984 numerous bag failures had begun to occur. The bags removed from the unit were very stiff, and were covered with a very heavy, semi-hard gray dust cake. Batt cracking was evident along the vertical cage wire lines. Permeability on these failed bags ranged from about 1.0 CFM/sq ft at 1/2" W.C. near the top of bags to a high of about 5.0 CFM/sq ft near the bottom of the bags. New bag permeability average was 35 CFM/sq ft. There was no significant recovery of permeability after heavy vacuuming. Mullen burst and batt peel tests indicated that 50 to 75 percent of the original strength was retained in the bags. The conclusion drawn from this data is that the primary mode of failure was flex fatigue caused by the "blinded" fabric. Obviously the upset moisture conditions experienced during the initial start up phase could have been a major

contributor to the fabric "blinding".

Bag failure experience and cost considerations early in 1984 led to the decision to rebag the unit with woven fiberglass material. Bag replacement was accomplished in March 1984 on the first compartment. After four of the six modules were rebagged, no improvement in the opacity was noted and in addition high pressure drops across the fabric filter had been experienced on several occasions. Deeper investigation of this situation revealed that a major problem appeared to be incorrect size (length) of the new bags.

At that point all new bags of the woven fiberglass construction were ordered. Bag manufacturing lead times did not allow the second rebagging to be completed until July, 1984.

Some high pressure drop problems were experienced shortly after bringing the fabric filter back on-line with all new woven glass bags in the first half of July. These problems were attributed to an improper cleaning cycle for the bags. After changing the cleaning cycle we were able to achieve a reasonable pressure drop across the fabric filter with all six modules on-line; however, cleaning frequency of the bags was greater than desired.

Bag failures began to appear in November 1984 as indicated by an increase in our background opacity. During September and October 1984 it was noted that the fabric filter could not be operated with only five modules on-line for any extended period of time. Even with maximum cleaning of the bags the pressure drop would build until after a few hours the pressure drop "run away" point was exceeded forcing the fabric filter to be bypassed. During this period of operation, pulse cleaning pressure had been increased to 100 PSIG, and cleaning frequency was at its maximum. It was also not possible to operate the fabric filter at normal load with the spray dryer off line, due to the increase in gas volume caused by the higher temperature entering the fabric filter.

Failed bags removed from the unit had numerous holes and obvious abrasion to the fabric, especially in the vertical cage wire areas. Permeability test results on the failed bags ranged from 0.5 to 3 CFM/sq ft (at 1/2" W.C.). Recovery after heavy vacuuming ranged from 3.5 to 12 CFM/sq ft. New bag permeability averaged 16 CFM/sq ft.

Mullen burst tests indicated that 65 to 75 percent of the original strength of the bag was retained after 5 months of operation. It was concluded that, due to the lower initial (new) permeability and poor recovery upon cleaning, the excessive cleaning frequency and high cleaning pressure resulted in premature flex failures.

In early February, 1985 rebagging was again completed on the unit. The bags installed consisted of a mixture of foreign and domestic acid resistant Nomex felt, Ryton on Ryton scrim and Ryton on Rastex (Goretex) scrim. The fabric data is summarized in Table 2.

TABLE 2. PHYSICAL PROPERTIES
OF
FELT BAGS INSTALLED FEB. 1, 1985

MODULE NO.	GENERAL			FINISH			SCRIM			
	FELT TYPE	BATT MAT'L.	WT.(OZ./YD. ²)	CHEMICAL	SURFACE	FIBER	YARN	WEAVE	COUNT(W x F)	WT.(OZ./YD. ²)
2	ACID RES. NOMEX [⊕]	NOMEX	15.0	ACID RES. (Zw 184)	SINGED (MOD.)	NOMEX	SPUN/PLIED	PLAIN	35 x 35	4.2
6	ACID RES. NOMEX [⊕] /GLASS	NOMEX/20% GLASS	13.4	ACID RES.	SINGED (LIGHT)	NOMEX	WOOLEN SPUN--	PLAIN	19 x 10	3.6
5	RYTON/RASTEX [⊗]	RYTON	15.0	NONE	NONE	RASTEX [⊗]	RIBBON	PLAIN	24 x 24	2.0
3	RYTON/RASTEX [⊗]	RYTON	13.6	NONE	NONE	RASTEX [⊗]	RIBBON	PLAIN	24 x 20	2.5
4	RYTON/RYTON	RYTON	18.0	NONE	NONE	RYTON	SPUN	PLAIN	19 x 14	3.1

⊕ HEIMBACH FELT
⊕ DOMESTIC FELT
⊗ FORMERLY GORE-TEX

FABRIC DATA SUPPLIED COURTESY OF
W. W. CRISWELL CO.

As of this writing all the fabrics installed in February 1985 have performed satisfactorily with no bag failures. More detailed information on the various fabrics and their physical condition to date is contained in the paper entitled, Evaluation of Acid-Resistant Synthetic Needled Felt Filter Media in Coal-Fired Boiler Pulse-Jet Baghouses by Mr. Theron Grubb which was also presented at this symposium.

Relative air flow measurements made on each of the six fabric filter compartments do not indicate any serious imbalance of flow among the various fabric types. Currently a pulse cleaning pressure of 60 PSIG is being used and a flange-to-flange pressure drop of 5 inches W.C. is being maintained across the unit.

ASH HOPPER HEATERS

Difficulty was experienced during the winter months of 1984 and 1985 in removing the ash from the fabric filter ash hoppers due to condensation in the hoppers. After reviewing the operating temperatures, heat loss, etc., it was decided to increase the electric hopper heating level from the existing 2.6 KW per module (117 watts per sq ft) to 4.4 KW per module (180 watts per sq ft). In addition, throat heaters were added to each hopper discharge. This modification was completed in early fall of 1985 and through January 1986 no problems have been experienced with hopper pluggage due to condensation.

ASH CONVEYOR

Initially the pneumatic ash conveying system was connected to the spray dryer bottom discharge as well as the fabric filter hoppers. However, due to early experience with wet ash at the spray dryer bottom, this connection was broken and the ash discharged into a hopper and later directly into a special truck. The normal moisture content of this ash over the last 18 months indicated that it could again be conveyed by the pneumatic ash system; however, this has not been attempted to date.

The ash system in general has functioned very well. There has been very little wear on the ash conveyor piping. The lime in fabric filter fly ash tends to coat the normal high wear areas and acts as a protective layer. On occasion this buildup has reached the point where manual removal was necessary. We have also experienced buildup in the steam eductor combining tube venturi, which requires chipping to remove.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

Session C-6: INTEGRATED CONTROL PROCESSES II

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RESULTS OF PILOT SCALE TESTS OF E-SOX

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ABSTRACT

Research aimed at developing a low-cost retrofit system, called E-SOX, for combined particulate and SO₂ removal is described. The E-SOX concept centers on recent advances in electrostatic precipitator (ESP) technology that make it possible to reduce the size of the ESP necessary to meet particulate emission limitations. Results of mathematical modeling and pilot plant experiments on the feasibility of using the freed space for SO₂ removal are presented. These results show that modest (40-65%) levels of SO₂ removal are possible in the freed space when lime is used as the reagent. High (90%+) levels of SO₂ removal are possible when sodium carbonate is used as the reagent. The results of data analysis to determine important process variables are also presented. This analysis shows that there is no correlation between SO₂ removal and calcium-to-sulfur ratio once the calcium-to-sulfur ratio exceeds about 1.5. An analysis of the particulate collection efficiency of the E-SOX system shows that an E-SOX system can meet a 45 ng/J particulate emission standard when the SCA of the particulate collection system is about 30 s/am.

RESULTS OF PILOT SCALE TESTS OF E-SOX

INTRODUCTION

Because of the concern over acid rain and the resulting interest in low cost methods for reducing sulfur dioxide (SO_2) emissions from existing plants, the Particulate Technology Branch (PTB) of EPA's Air and Energy Engineering Research Laboratory (AEERL) at Research Triangle Park, NC, has begun work to combine advanced ESP and SO_2 removal technologies in an existing ESP. This process, called E-SOX, combines advanced ESP technology and spray dryer technology to remove particulate matter and SO_2 in the same physical equipment.

The original analysis of E-SOX was presented by Sparks et al.(1) The original analysis showed that E-SOX was technically feasible and that the process economics were favorable when lime was used as the SO_2 sorbent with stoichiometry of 2 or less. The work in this paper covers additional experiments of SO_2 and particulate removal and additional analysis of the data from the previous experiments.

DESCRIPTION OF PILOT PLANT

The experiments were conducted in the ESP pilot plant located at AEERL. The pilot plant is a four-section single-lane pilot plant capable of operating from ambient to 350°C . Sampling ports are located at the ESP inlet, the ESP outlet, and between each pair of electrical sections. Gas for the pilot plant is provided by heating outside air with natural gas burners.

The configuration for these experiments is shown in Figure 1. Sections 1 and 2 were converted to SO_2 reaction chambers by removing the discharge wires. The SO_2 was injected at point A located before the gas distribution plate. Scrubbing slurry was injected at point B using sonic spray nozzles. These nozzles are two fluid nozzles and produce drops in the range of 5 to $10\ \mu\text{m}$ in diameter according to the manufacturer. SO_2 was measured at four locations (C, D, E, and F). Location C provided about 0.5 sec of residence time; location D, 1.3 sec; location E, 2.1 sec; and location F, 3.7 sec.

For the E-SOX tests the pilot plant was operated with a gas inlet temperature (measured before the spray nozzles) of 150°C and a gas flow rate of about $0.487\ \text{am}^3$ at 150°C . The ESP was operated under about 5 cm H_2O negative pressure to ensure that no SO_2 was released into the room. SO_2 and air flow rates were measured to determine the air leakage into the ESP. These measurements showed about 5% leakage between the inlet and the sampling ports at the end of section 2. There was 20% leakage between the ESP inlet and the ESP outlet. The E-SOX performance data were corrected for this leakage.

SO₂ was measured with a TECO analyzer that was zeroed and spanned periodically. The SO₂ injection rate was between 1100 and 1500 ppm for most tests. A few runs were made with SO₂ injection rates as low as 500 ppm. Some preliminary experiments were run with SO₂ concentrations in excess of 2,500 ppm.

Two sets of SO₂ experiments were conducted: sodium carbonate was used as the scrubbing reagent for the first set; and lime was used for the second. The experimental procedure was the same for both sets of experiments.

Particulate collection experiments were conducted by converting the ESP into a multistage ESP. Cooled pipe prechargers were installed in sections 1 and 3, and 0.95 cm diameter discharge electrodes were installed in sections 2 and 4. In this configuration the ESP had a specific collector area of 12.6 s/am.

Particulate measurements consisted of inlet and outlet particle mass and particle size distribution. Particle size distributions were determined with Meteorology Research, Inc. (MRI) cascade impactors. The impactors were calibrated by AEERL personnel. Impactor data reduction was performed using procedures developed by AEERL and Denver Research Institute.

RESULTS

SO₂ Experiments

A number of SO₂ removal experiments were conducted to determine the efficiency of the E-SOX system. The experiments were also designed to determine the importance of process variables such as residence time, liquid injection rate, stoichiometry, and sorbent concentration. The experiments were conducted with sodium carbonate and lime as the sorbents. As is discussed below, the two sorbents behaved quite differently.

The data for sodium carbonate are summarized in Figures 2 and 3. Figure 2 shows that with sodium carbonate the initial SO₂ removal rate is quite high and that SO₂ removal continues, but at a lower rate, throughout the system. The initial high removal rate corresponds to the time that liquid drops exist in the system. During this initial time, SO₂ removal occurs due to mass transfer to liquid drops and reaction with sodium carbonate in liquid solution.

During the second phase SO₂ occurs due to mass transfer to dry sodium carbonate. The rate of SO₂ removal is much lower in this second phase.

Figure 3 shows that SO₂ removal efficiency is a function of stoichiometry for sodium carbonate. The higher the Na/S ratio, the higher the SO₂ removal efficiency.

The results for lime are shown in Figures 4 and 5. The lime data show moderate SO₂ removal rates for the initial fraction of a second similar to that obtained with sodium carbonate. As before, this initial rapid rate of SO₂ removal is governed by mass transfer to liquid drops and reaction of SO₂ with the lime contained in the drops. Note, however, that the lime is not as efficient in removing SO₂ in this initial phase as was the sodium carbonate. This suggests that liquid phase reactions between lime and SO₂ are important rate limiting steps.

The data also show that for lime under the conditions of these experiments (approach to saturation greater than 30°C) there is little or no reaction between the lime and the SO₂ after the drops have evaporated.

Figure 5 shows that once the Ca/S ratio exceeds about 1.5 to 2 there is no benefit to increased stoichiometry. This result is in agreement with data reported by Bechtel (2) for a short residence time spray dryer.

Figure 6 shows that there is a significant benefit to high slurry injection rates.

Particulate Experiments

Limited experiments were conducted to determine the ability of the multistage ESP concept to operate under E-SOX conditions. The experiments were conducted with a gas flow rate through the ESP of 0.4744 am³/s (am³ is actual cubic meters). The dust used in the experiments was generated by injecting fly ash and lime slurry.

The results for penetration as a function of particle diameter are shown in Figure 7. The overall mass efficiency for all the experiments ranged from 98 to 99.5%.

DISCUSSION

SO₂

The experimental data demonstrate the technical feasibility of the E-SOX process. Moderate SO₂ removal can be achieved in the short residence times available in the first electrical section of an ESP.

The lime data show that lime injection rates should be limited to that necessary to produce a stoichiometry of 1 to 1.5 instead of 2 to 3 as originally thought. Lime injection at these stoichiometries will improve the economics of the process by reducing the capital costs associated with the lime handling system and by reducing the operating costs associated with sorbent costs. The low lime stoichiometry will also result in lower particulate loading for the ESP than originally thought.

The data indicate that liquid-phase reactions, probably the rate of lime dissolution, play a significant role in limiting SO₂ removal rates in the liquid drop phase of the E-SOX process. Improved SO₂ performance could be obtained by finding additives that increase the solubility of lime in the solution.

Significant increases in SO₂ removal are also possible if the reactivity of the dry lime can be increased. Research on sorbents for spray dryers at AEERL indicates that sorbents can be found that will allow significant SO₂ removal after the drops have evaporated. The sorbents developed under this program will be evaluated for E-SOX as part of future E-SOX research.

Particulate

The particulate data were used to estimate the emissions from a full scale ESP with the E-SOX retrofit. The estimation was performed by first converting the penetration vs particle diameter results to effective migration velocity vs particle diameter as shown in Figure 8.

These effective migration velocities were then used to estimate the overall penetration of a full scale ESP collecting fly ash and lime from the E-SOX process. The estimation was obtained by integrating the following equation over the particle size distribution:

$$Pt_0 = \exp -w(d) SCA \int f(d) dd$$

where Pt_0 is the overall mass penetration, $w(d)$ is the effective migration velocity of particles of diameter d , SCA is the SCA of the ESP, and $f(d)$ is the fraction of particles with diameters between d and $d + dd$. The result given by the above equation was then corrected for sneakeage, non-uniform gas flow distribution, and rapping reentrainment. (In theory the above corrections should not be needed since the effective migration velocity should include the effects of non-ideal factors. However, since the sneakeage and the rapping reentrainment in the pilot ESP are lower than the sneakeage and rapping reentrainment in full scale ESPs, the corrections were included.)

The lime particle size distribution was based on impactor data obtained as part of the E-SOX experiments. The fly ash particle size distribution was assumed to be log normal with a mass mean diameter of 15 μ m and a geometric standard deviation of 3.5. This is consistent with field measurements of fly ash size distributions for high sulfur coals.

The results of these estimates are shown in Figure 9 in terms of emissions versus SCA. Note that, if the emission standard is 45 ng/J, a common standard for older plants, an SCA of at least 28 s/am is required. This means that an ESP with a before-retrofit SCA of 41 s/am could be fitted with E-SOX and still meet required particulate emission limits.

The ESP modeling did show that non-ideal factors are a major cause of particulate emissions from the E-SOX process. Thus, in order to obtain compliance, sneackage, gas flow non-uniformities, and rapping reentrainment must be minimized. Rapping reentrainment may be especially severe in the E-SOX ESP, and research on ways to minimize the impact of rapping reentrainment will be conducted in the next several months.

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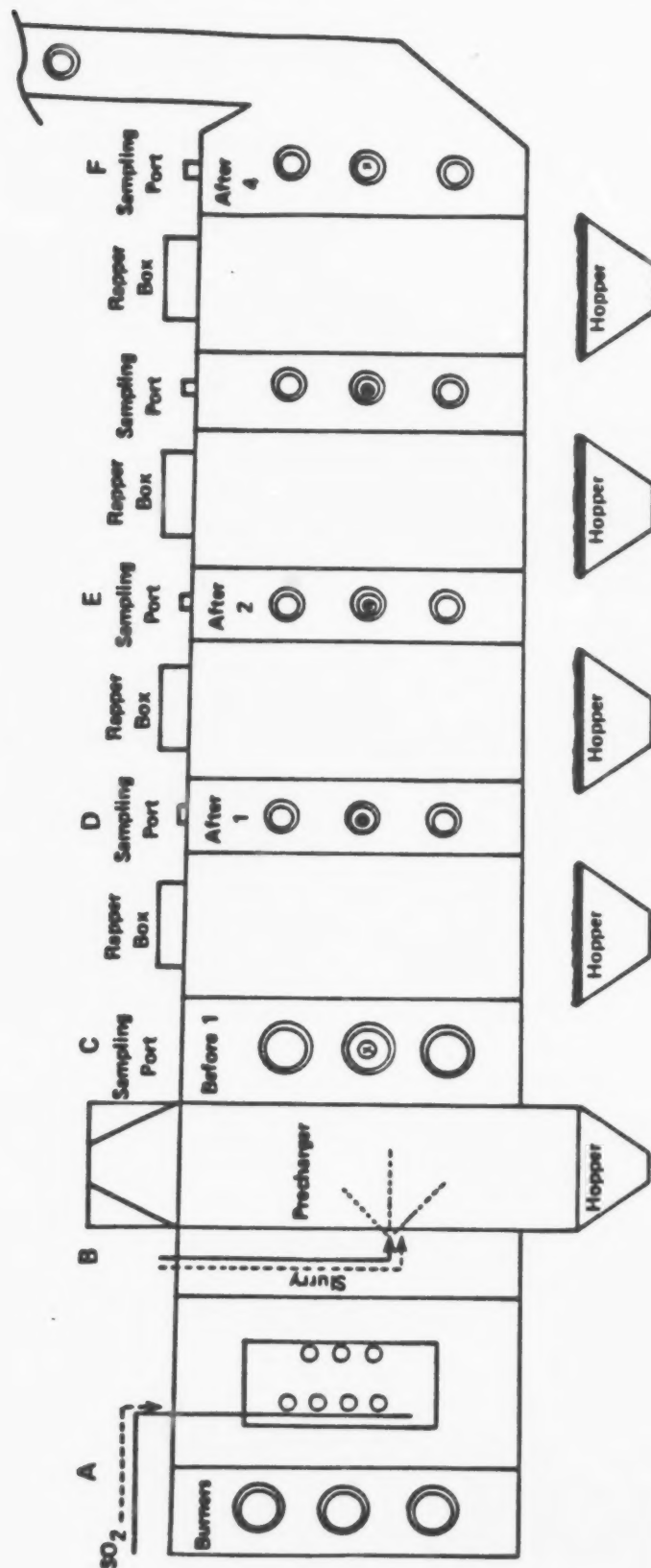


Figure 1. Sketch of pilot plant modified for E-SOX

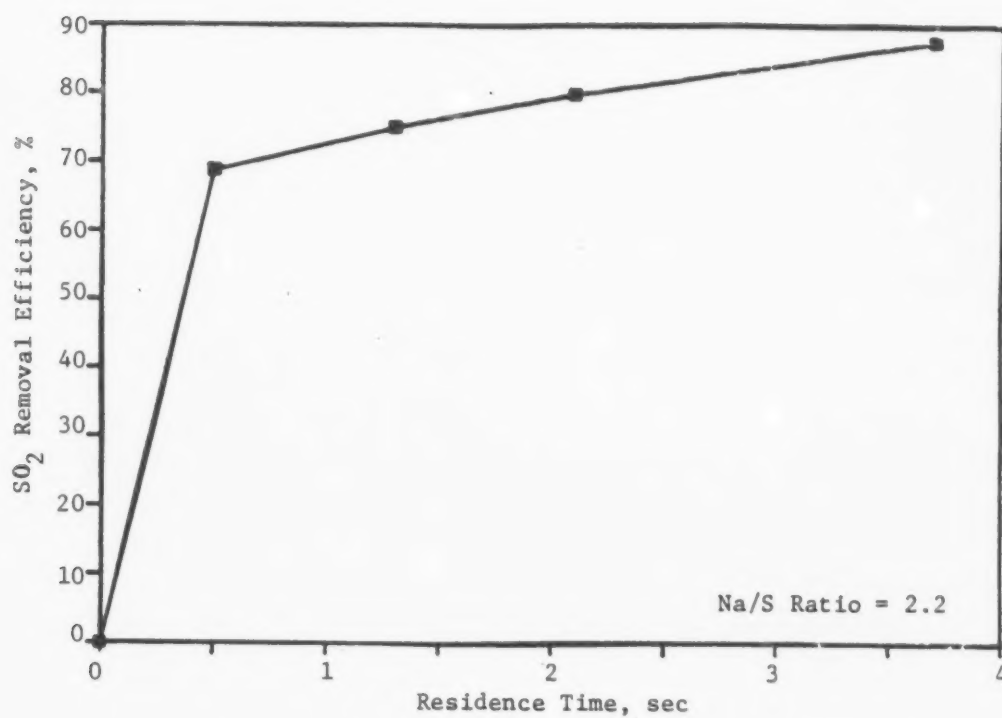


Figure 2. Sodium carbonate SO₂ removal efficiency.

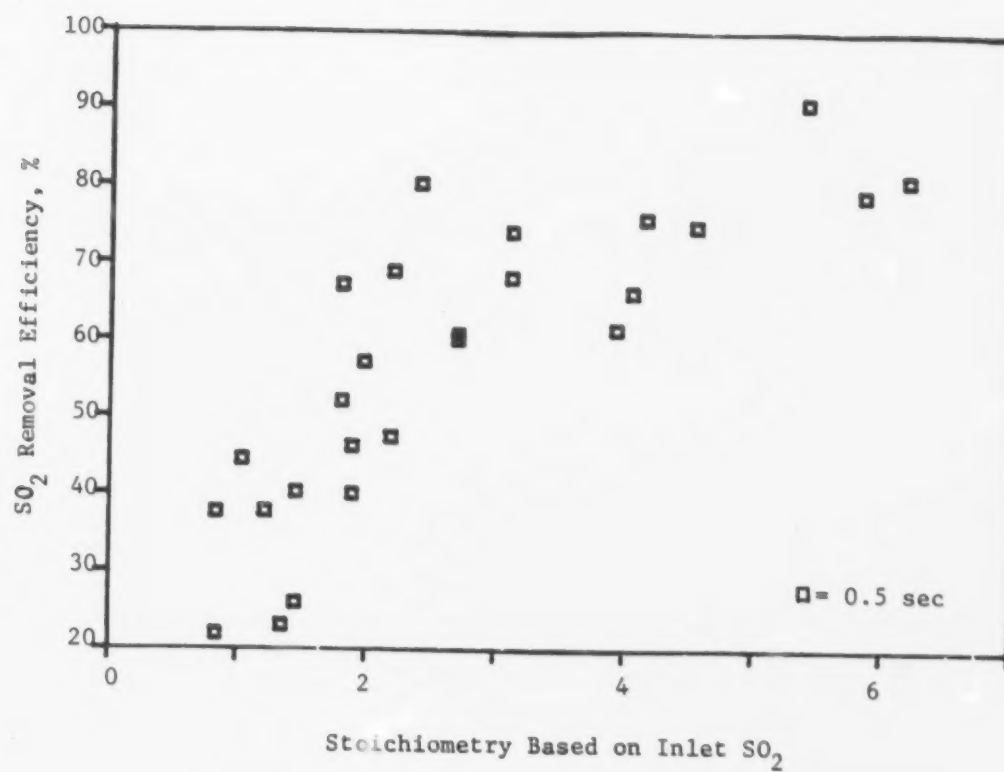


Figure 3. Sodium carbonate effects of stoichiometry.

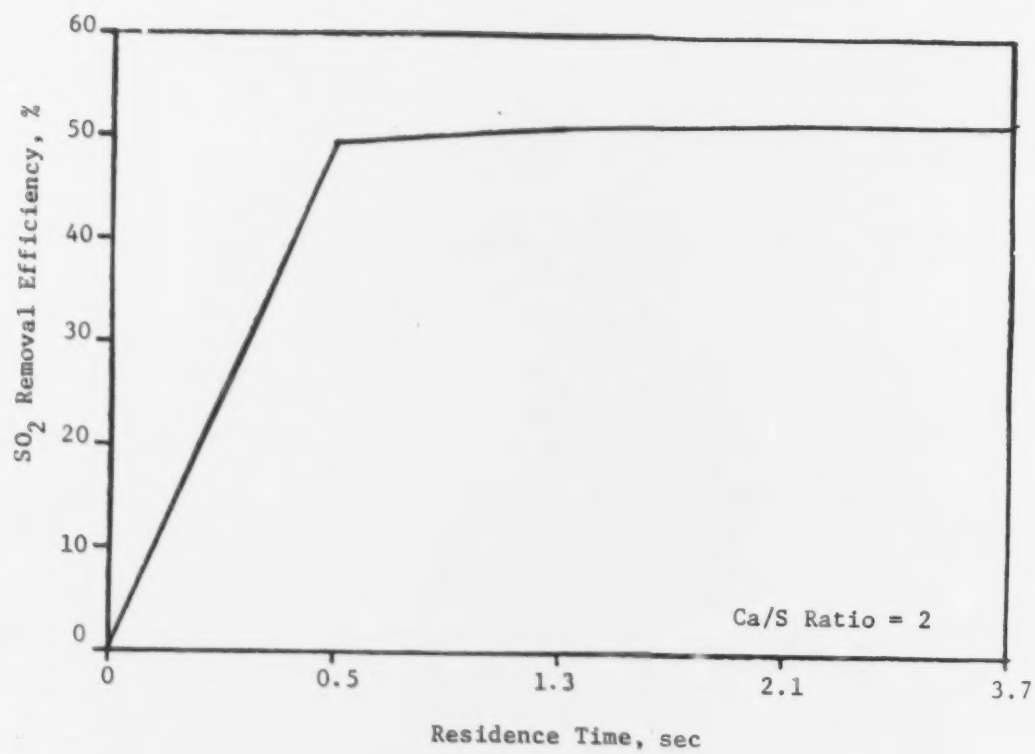


Figure 4. Lime SO₂ removal efficiency.

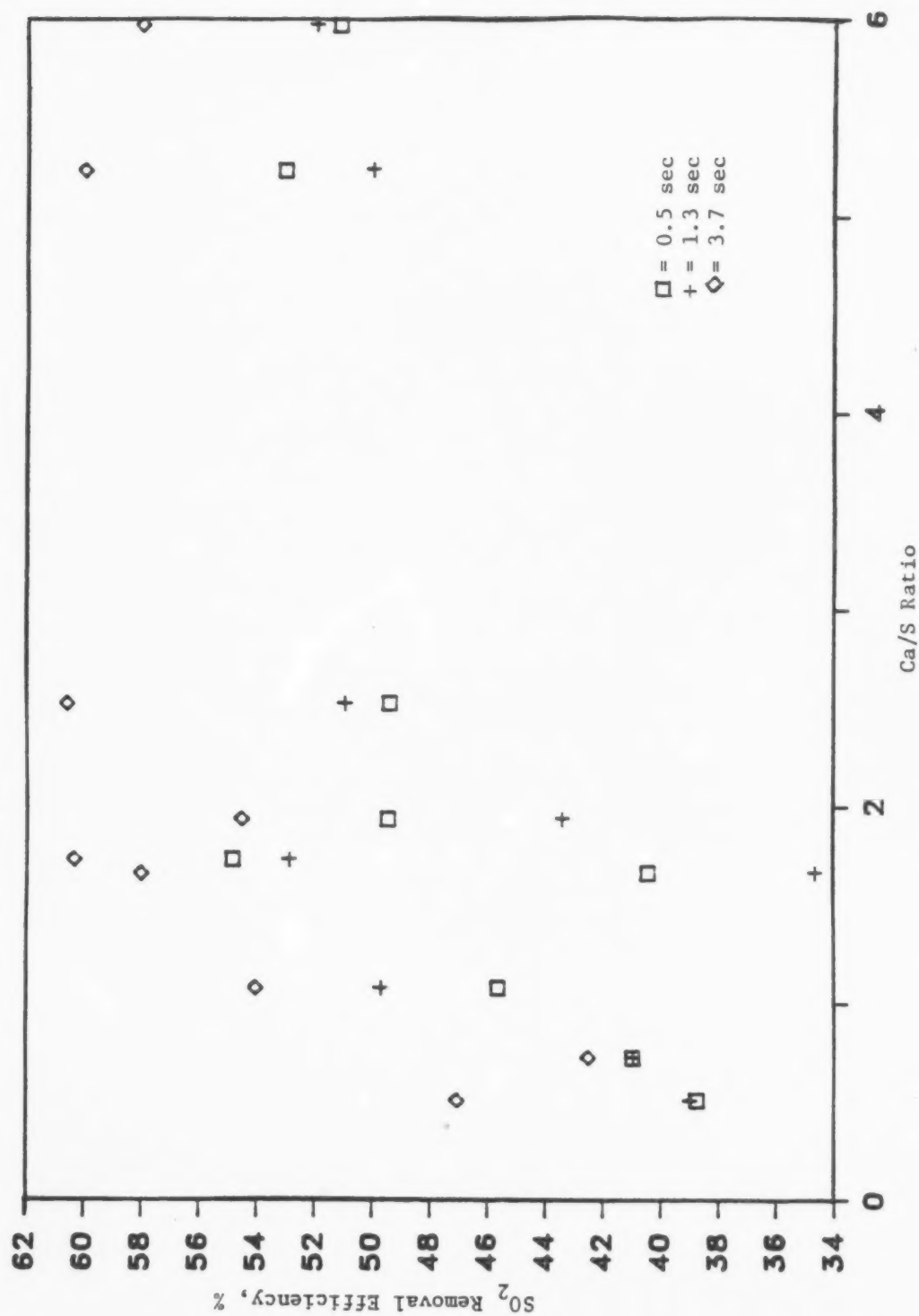


Figure 5. Lime effects of Ca/S ratio for all residence times.

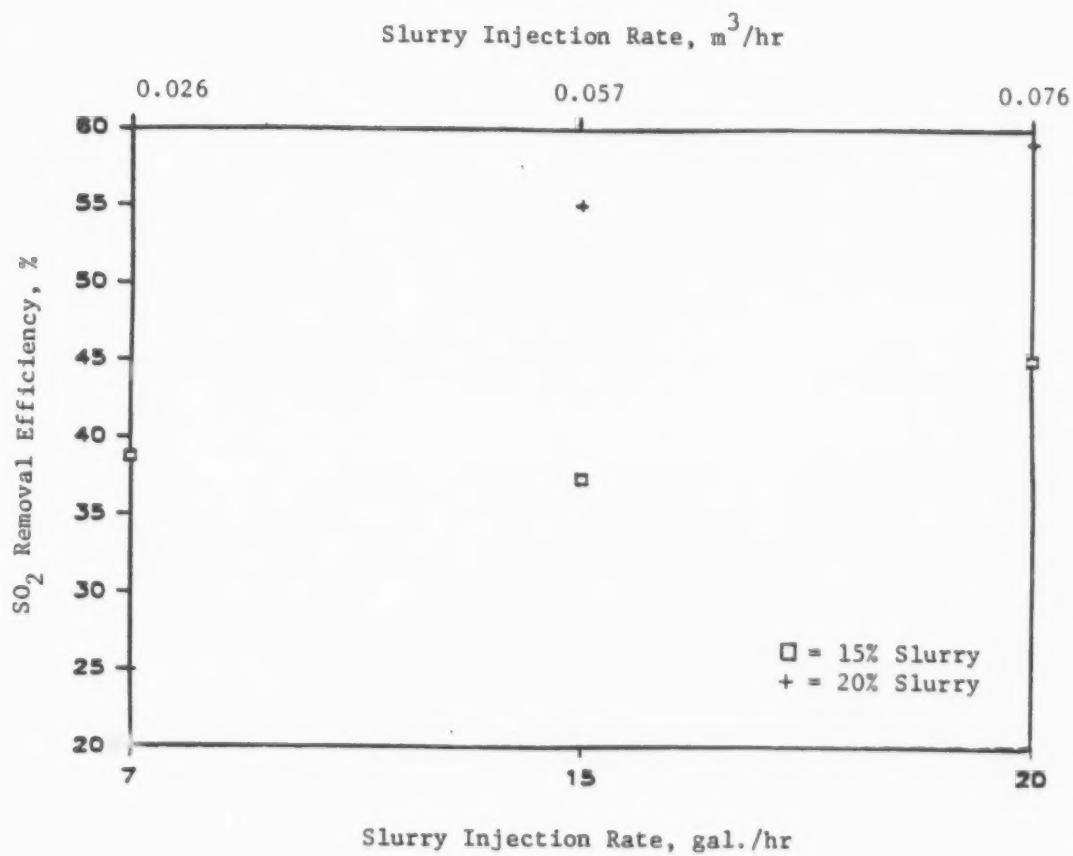


Figure 6. Lime data for E-SOX showing effects of injection rate.

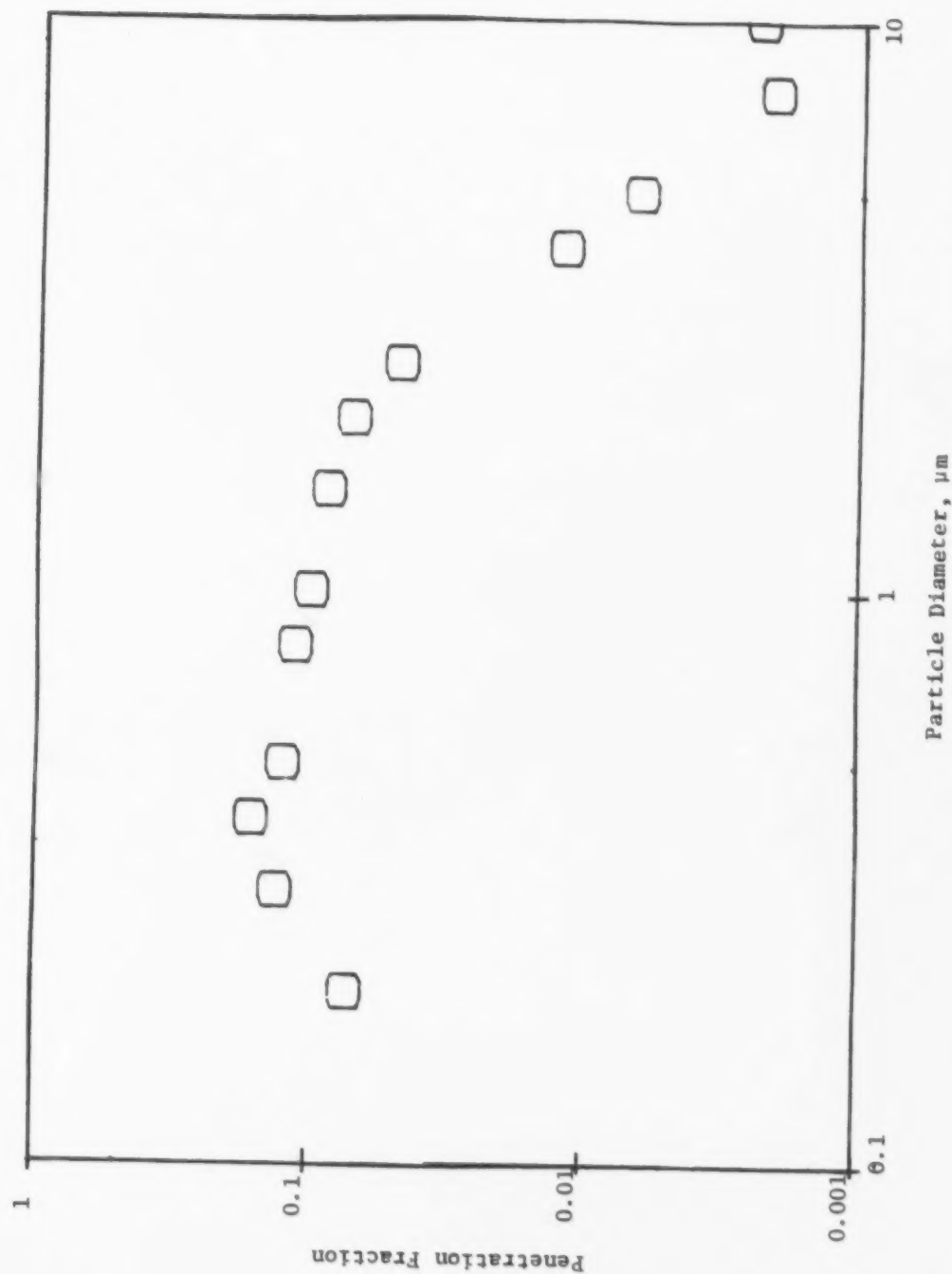


Figure 7. Penetration as a function of particle diameter.

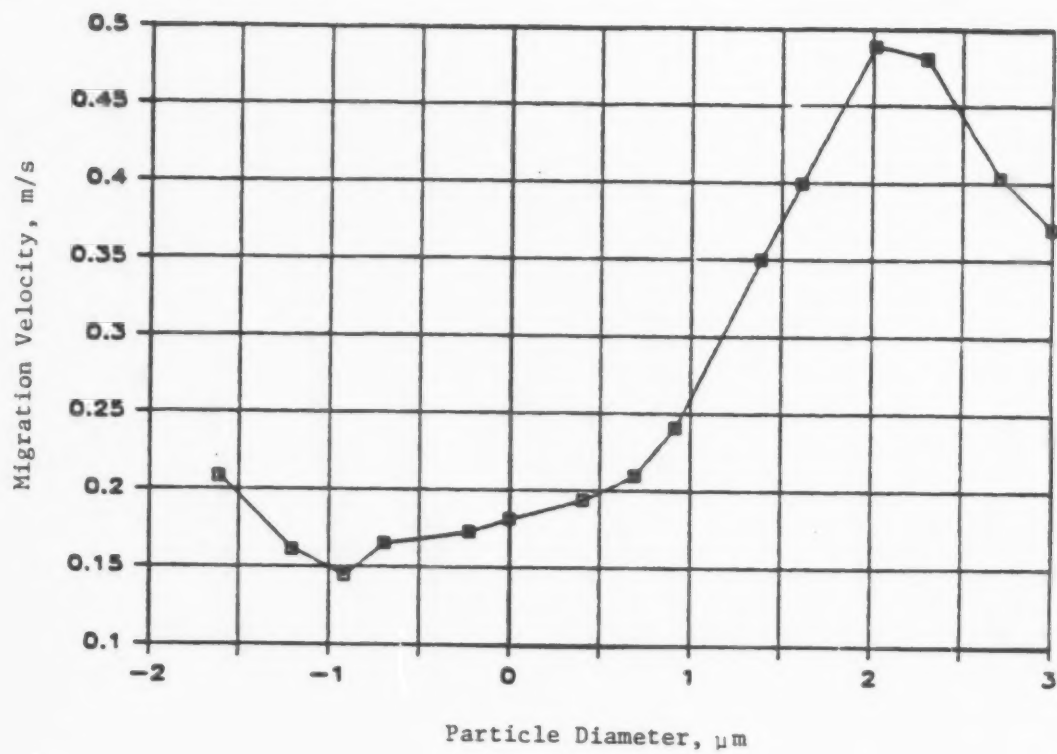


Figure 8. Migration velocity vs particle diameter.

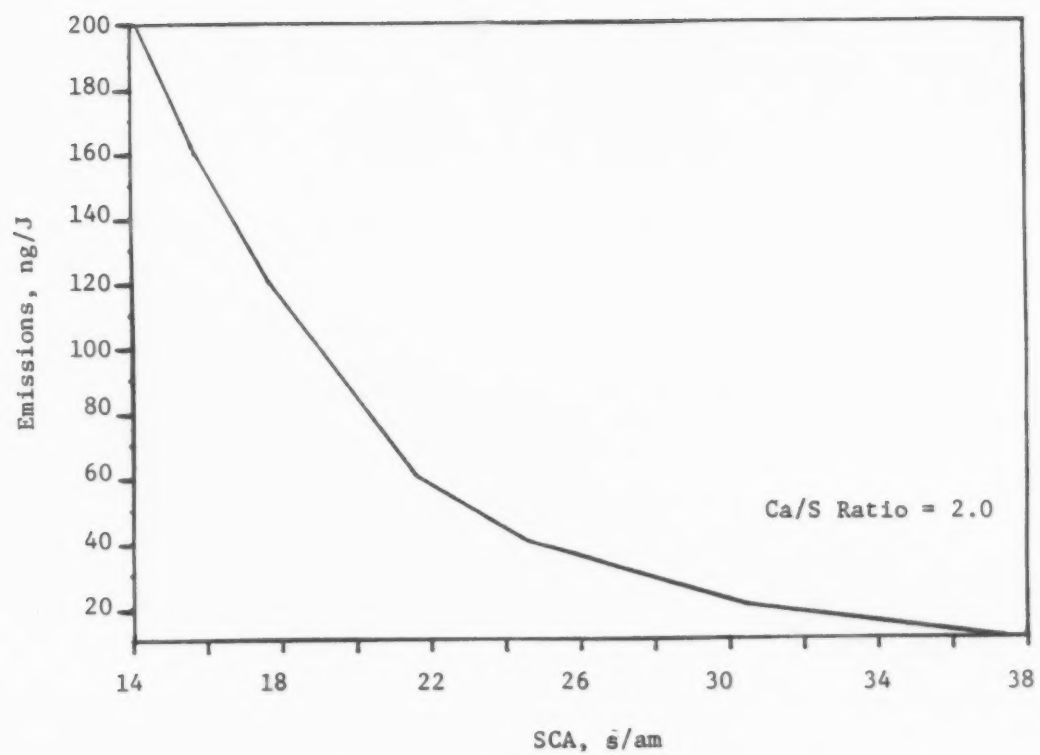


Figure 9. E-SOX particulate emissions.

RESEARCHES FOR OPTIMAL INTEGRATED FLUE GAS TREATMENT

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SUMMARY

The current status of pollutant abatement systems installed in power stations in Italy is presented and the future developments for facing the need of more energy from coal and for matching with more stringent air quality standards are discussed. A careful planning for optimal integrated pollutant control is supposed necessary for which reliable information on current available technologies and on developing new ones are needed.

ESP are assumed to remain, at least in the near future, the most favourable choice for particulate abatement especially if new technologies as pulse energization and pre-charging will confirm their benefits; special attention will also be devoted to available and developing DeNO_x and DeSO_x processes.

The paper presents some results of the several research projects which are conducted in Italy in order to assess the performance parameters of the different technologies for the control of the emissions from thermal power plants and to provide the technical and economical information needed for an accurate planning both for retrofit applications and for new plants.

INTRODUCTION

The first thermoelectric plants, sized 35 MWe, appear in Italy during the years '50s and the only device for atmospheric pollution control is a mechanical collector.

At the beginning of the years '60s, the unit size increases to 160 - 240 MWe and the units burning pulverized coal are equipped with electrostatic precipitators.

During the years '70s the unit size increases to 320 MWe and the fuel is generally oil; the electrostatic precipitator is not compulsory, the amount of solid particle emissions being well below the safeguard limit, but it is generally installed for facing specific operating conditions as start-ups, soot-blowings etc.

At the end of the years '70s, as a consequence of the oil crisis which made more economically convenient the use of coal and of an increased attention to the protection of the environment not only the new units are planned with an electrostatic precipitator of increased performances but also an effort is made for upgrading the electrostatic precipitators installed in existing units.

A research project started at the beginning of the years '80s with the following main objectives:

- to increase ESP reliability;
- to decrease the electrical energy consumption of ESP;
- to increase ESP flexibility, i.e. the ability to maintain constant performances with different operating characteristics (specifically with different coals);
- to increase the collecting efficiency of the existing units;
- to decrease the size and to reach a better cost/benefit ratio in new equipment.

Meanwhile the increased attention to air pollution suggested to investigate on the possibility to decrease the sulphur and nitrogen oxides emissions, both acting on fuel quality and on combustion process and developing criteria for optimal integrated flue gas treatment.

The technology employed in Japan for NO_x abatement foresees catalytic processes on the flue gas upstream of the air pre-heater; this entails a hot-side precipitator while in the existing plants a cold-side ESP is installed.

The abatement of SO_x is most frequently achieved with complex chemical treatments which are quite expensive and give by-products whose disposal may rise management problems.

It is then justified an effort for verifying the feasibility of alternate techniques and a careful assessment of the costs and benefits of the different means for the abatement of the flue gas pollutants.

In this paper a report is given on the research work undertaken, on the results obtained up to now and on the plans for future activity.

PULSE ENERGIZATION OF AN INDUSTRIAL ESP

Since the fundamental studies of H. White [1], S. Masuda [2] and several others investigators, the influence of dust resistivity on the precipitator efficiency through the back-corona phenomenon is well

recognized as one of the principal sources of poor precipitator efficiency.

The means for facing the back-corona may be divided into two broad groups: a) chemical conditioning of the flue gas or thermal conditioning of the electrode in order to decrease the resistivity of the dust layer and b) careful control of the electrical process through the parameters of the electrical energization.

The SO₃ conditioning of the flue gas seems to be the most attractive mean out of the first group; the use of pulse voltage waveforms seems to be the most attractive mean out of the second group. The latter, appearing, at least in principle, less expensive and of easier use than the former, was assumed as subject of investigation.

A prototype of pulse power supply, the main characteristics of which are reported in Table I, was installed in a 320 MWe ESP (fig.1) the main characteristics of which are summarized in Table II.

TABLE I
Characteristics of the pulse power supply used during the tests

D.C. base voltage level	up to 50 kV
pulse voltage amplitude	up to 50 kV
pulse repetition rate	10 to 400 Hz
pulse duration	60 microseconds

TABLE II
Characteristics of the ESP used for the experimental campaigns

flue gas flow rate	Nm ³ /h	1,090,000
temperature	°C	140
inlet dust load	g/Nm ³	10
efficiency	%	99.60
number of parallel channels . . .	n.	4
number of electrical fields . . .	n.	4
number of gas passages	n.	128
width of gas passage	mm	228.6
height of collecting plates . . .	m	10.06
length of field: first	m	1.83
second to forth	m	2.74
gas velocity	m/s	1.56
collecting area: first field . . .	m ²	1,177
second to forth field	m ²	1,766
total collecting area	m ²	25,900
specific collecting area	m ² /(m ³ /s)	56.5

TABLE III
Summary of the 83 and 84 test campaigns

campaign	first	second	third
number of tests	14	21	31
test channel	B	B	A
reference channel	C	C	C
fields energized: test channel	3B	3B	1A-2A
reference channel	3C	3C	1C-2C
impulse energization parameter			
DC base voltage (KV)	22	22	20-22
impulse voltage (KV)	35-30	25-30	25-30
pulse duration (10^{-6} s)	60	60	60
pulse rate Hz	400	400	400
conventional voltage (KV)	30	30	30
outlet dust loads (g/Nm^3)			
with conventional energization	0.520 ± 0.139	1.675 ± 0.257	1.393 ± 0.526
with pulse energization	0.347 ± 0.077	1.158 ± 0.240	0.791 ± 0.096

The tests were always essentially based on the output dust load measurements of a test channel, which could be differently energized, and of a reference channel. The operating conditions were considered through the measurement of several quantities generally supposed influencing the precipitator efficiency and collectively by means of the output emission of the reference channel [3].

During 1983 and 1984 three test campaigns were carried out by pulse energizing either one or two fields. The main characteristics and results of these three campaigns have been already reported [4]; they are summarized in Table III with reference to the precipitator layout of fig. 1.

In 1985 two more extensive test campaigns have been carried out; the first lasting five weeks, the second lasting about two weeks. During these tests a computer was used for controlling the pulse energization, different qualities of coal were supplied to the boiler and a larger number of operating and influencing parameters was measured.

For each test the following quantities were measured:

- Dust load concentration at the input and output sections of the test and reference channel, following ASME specifications;
- Boiler operating parameters: electrical load, injected SO_3 (when applicable), flue gas temperature;
- Resistivity of the dust following the in-situ point-plate method;

- Electrical characteristics of the power supplies: number of sparks, voltage and current values, power and energy consumptions;
- Coal characteristics: ash, sulphur, hydrogen and nitrogen content; humidity, lower specific heating value;
- Dust characteristics: chemical analysis of dust, particle size distribution.

Five configurations of the test channel were investigated:

- a : the four fields conventionally energized
- b : the four fields pulse energized
- c : the first field conventional,
the second to the forth field pulse energized
- d : as "c" with SO₃ conditioning
- e : as "a" with SO₃ conditioning
- f : the first and the second field conventional,
the third and the forth pulse energized

TABLE IV
Main results of '85 test campaign

A	B	C	D	E	F	G
1 - 11	a	#1	$(1.3-16) \times 10^{12}$	14.673 $\pm$ 3.950	2.375 $\pm$.359	83.81%
12 - 20	b	#2	$(1.0-3.5) \times 10^{12}$	15.314 $\pm$ 3.103	.799 $\pm$.297	94.78%
21 - 26	c	#3	$(1.3-3.9) \times 10^{12}$	14.206 $\pm$ 4.921	.641 $\pm$.218	95.49%
27 - 28	d	#3	4×10^{11}	16.810 $\pm$ 5.459	.358 $\pm$.223	97.87%
29 - 33	e	#4	$(2.2-3.7) \times 10^{11}$	9.235 $\pm$.674	.080 $\pm$.040	99.13%
34 - 36	e	#5	$(1-2) \times 10^{10}$	13.380 $\pm$ 3.208	.076 $\pm$.034	99.43%
37 - 46	f	#2	$(8-20) \times 10^{11}$	14.707 $\pm$.875	.427 $\pm$.079	97.10%
47 - 52	a	#2	$(2.5-8.6) \times 10^{11}$	14.190 $\pm$.706	.982 $\pm$.140	93.08%
53 - 54	e	#2	5×10^{09}	13.505 $\pm$.207	.109 $\pm$.004	99.19%
55 - 56	e	#6	1.1×10^{10}	10.273 $\pm$ 1.618	.088 $\pm$.004	99.14%

A : test run n.

B : test set-up (see text)

C : type of coal (see aside)

D : in situ dust resistivity (ohm cm)

E : inlet dust load (g/Nm³)

F : outlet dust load (g/Nm³)

G : average efficiency (%)

#1 : Mix Australian and

South African/ Shell coal

#2 : South African / TCOA coal

#3 : South African/Shell

#4 : American-ash

#5 : South African BP

#6 : South African/Gen. Mining

The main results of these two campaigns, summarized in Table IV, leads some first remarks.

The first campaign started with a poor precipitator performance (about 84% efficiency) which was raised to about 95% efficiency with the pulse energization and finally to 99.4 % with SO₃ conditioning.

The second campaign, carried out after a maintenance and cleaning of the precipitator, shows a higher efficiency with conventional energization (in the order of 93%). In that case the pulse energization of the last two fields raised the efficiency to 97%. It is worth noting in evaluating the performances with the pulse energization that when two or more fields were pulse energized a single pulse supply was used, thus reducing drastically the sectionalization of the ESP.

During the first campaign the pulse energization of the four fields raised the w_k value of a factor 2.7 which is equivalent to an increase of the SCA over 60%; during the second campaign the pulse energization of the last two fields raised the w_k value of a factor 1.74 which is equivalent to an increase of the SCA over 30%.

TABLE V
Economical comparison

option (see text)	a)	b)	c)
capital actualized cost	100	72	83
operating actualized cost	16	32	4
total actualized cost	116	104	87

The energy consumption was particularly focused during the tests 35 to 56. As it is shown in fig. 2, the electrical power decreases when using the pulse energization and increases when the SO₃ is injected for controlling the dust resistivity.

A tentative economical comparison between the the following three options: a) a conventional ESP; b) an ESP with SO₃ conditioning; c) an ESP with pulse energization gave the results outlined in Table V. Several hypothesis have been made for estimating the costs for the different solutions and particularly the following:

- the pulse energization will almost double the e.m.v.;
- the SO₃ conditioning will increase the e.m.v. by a factor 2.5;
- the required efficiency is 99.8%.

OPTIMAL PULSE POWER SUPPLY CONTROL AND ELECTRODE GEOMETRY

The achieved investigations suggested that the pulse power supply control may be of great importance in assuring a permanent highest

efficiency. Some theoretical considerations suggest to correctly set first the DC base voltage, second the impulse voltage and finally control the frequency of the pulses; control devices available on the market simply assume a constant ratio between crest and mean value of the voltage and controls first the pulse frequency and second the voltage value. The experience from the tests carried out seems to suggest that once the mean voltage value (D.C. base voltage) and frequency of pulses are correctly set, the most sensitive parameter is the crest value of the pulses.

The electrode geometry, i.e.: the gas passage width, the emitting wire diameter and spacing, may also influence the collecting efficiency with pulsing voltage waveforms in a different way they do with conventional energization.

For assessing some general guidance criteria in the design of pulse energized ESP, further investigations are planned at a 35 MWe coal unit equipped with an ESP (PRISMA) of specific design to incorporate two independent channels: a standard reference channel and an experimental channel in which the gas passage width can be varied from 300 mm to 500 mm and which can be energized alternatively with conventional and pulse generating power sets. The design data are given in Table VI, and fig. 3 gives a picture of the device. All the experiments will be monitored and controlled by a computer following different logic procedures.

Table VI
Main characteristics of the test facilities

unit		PRISMA(1)	IPSE(2)	CORFU(3)
Equivalent power	MWe	35	1.25	0.2
Gas flow	m ³ /s	75	2.70	0.5
Gas passage width	mm	300-500	300-500	300-600
Collection plate height	m	13	3.4	1.5
Aspect ratio		1.15	2.9	5.0
Treatment time	s	21	14	7.5
Electric fields	n.	3	2	5

(1) PRISMA : Precipitatore Industriale Sperimentale di Marghera

(2) IPSE : Impianto Pilota Studi sulla precipitazione
Elettrostatica

(3) CORFU : Condotto per Ricerche sui Fumi

Aside the 35 MWe industrial unit, a pilot unit of 1.25 MWe equivalent power (IPSE) will be installed to perform preliminary investigations without penalizing the electric production of the plant. In Table VI the main design data of this unit are also given while fig.4 shows a sketch of the device. This experimental pilot plant is also fully

monitored; particular attention has been placed in continuously monitoring the input and output dust loads and the fractional collecting efficiency along the precipitator length.

The research program is complemented by a 0.2 MWe loop (CORFU) to be installed at an experimental facility site. The 0.2 MWe loop has been conceived for testing advanced energization techniques, pre-chargers, and improved mechanical design and for calibrating both standard and more sophisticated measuring devices.

ADVANCED DESOX AND DENOX TECHNIQUES

The electron beam desulfurization and denitrization process received a careful attention, principally by the Japanese scientists; in 1981 Prof. Masuda released the final report of a research project on this subject sponsored by the Japanese Ministry of Education [5]. The difficulty in applying industrially this process seems to be principally related to the installation of an electron beam gun.

It is known that corona discharge under step rising impulse voltages develops with a front where high energy electrons are produced while the tail is filled with lower energy electrons; experimental investigations and theoretical studies suggest a Druyvenstein distribution for the energy of the electrons produced by a positive point-plate impulse corona discharge like that of fig. 5 [6]. It seems thus possible to produce the electrons able to start the ionizing processes through convenient corona discharges in suitable electrode arrangements. The results of some preliminary laboratory tests confirm that hypothesis [7].

The favourable aspects of such a technique, at least when burning low sulphur coal, are so remarkable to suggest further researches. An experiment has been planned with real flue gas using a test cell having a gas flow of about 1000 m³/h.

Different electrode geometries are supposed to be tested in the reactor module with voltage pulses having a rise time in the range of 100 ns and a crest value in the range of 100 kV. Tests will be carried out at different flue gas temperatures, with and without the injections of chemical reagents.

A computer model is also available for the chemical reactions triggered by the radiation [8]; an overall computer model is under development which should fit this program with that for corona discharge for providing criteria useful in designing a pilot plant.

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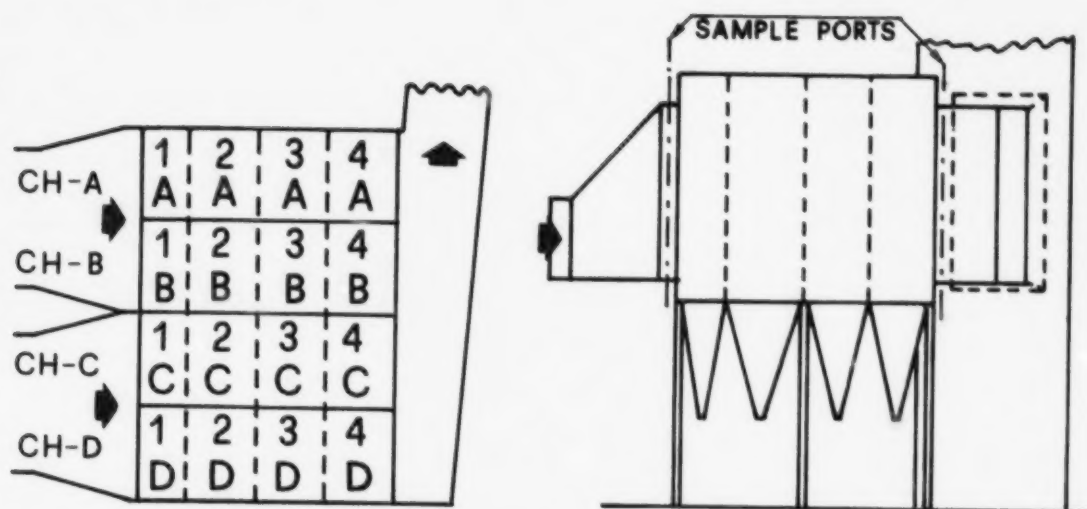


Fig. 1 : Sketch of the ESP downstream of a 320 MWe pulverized coal boiler where the experiments have been carried out.

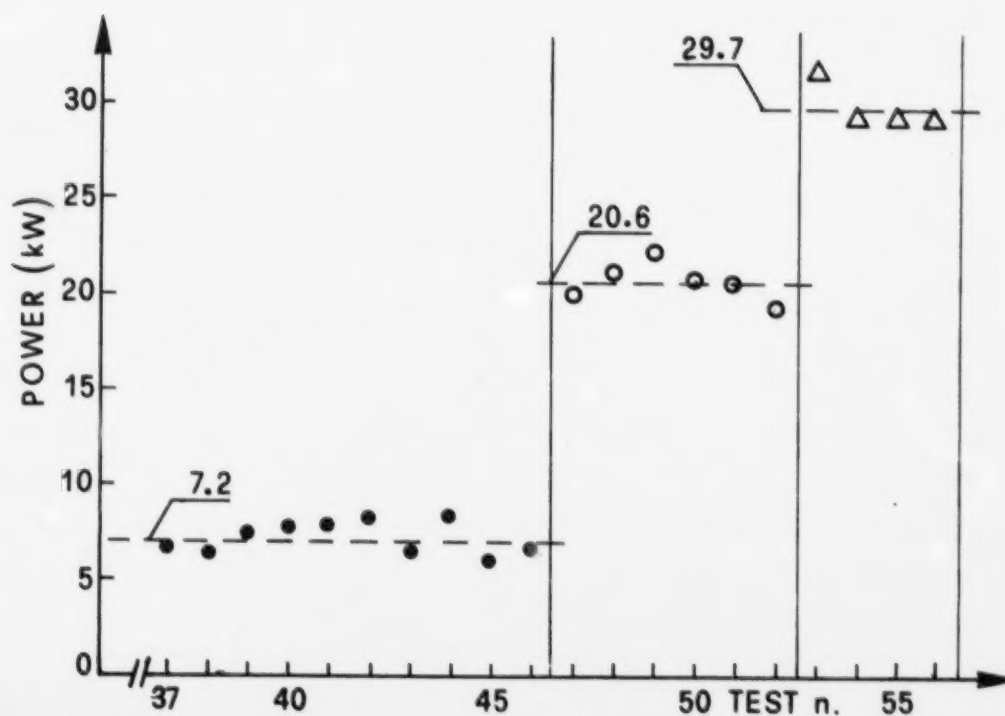


Fig. 2 : Power for the energization of the two last fields of the test channel. Tests 37 to 46 refers to pulse energization, tests 47 to 52 to conventional energization and tests 53 to 56 to conventional energization and SO_3 conditioning.

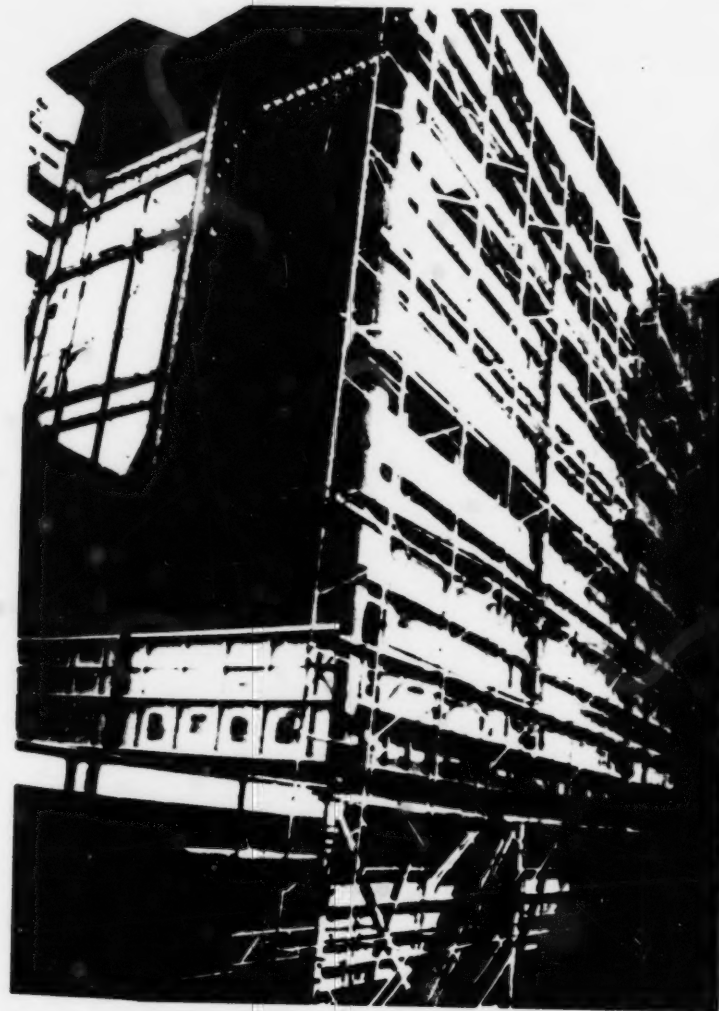


Fig. 3 : Picture of the experimental ESP
(PRISMA) under construction.

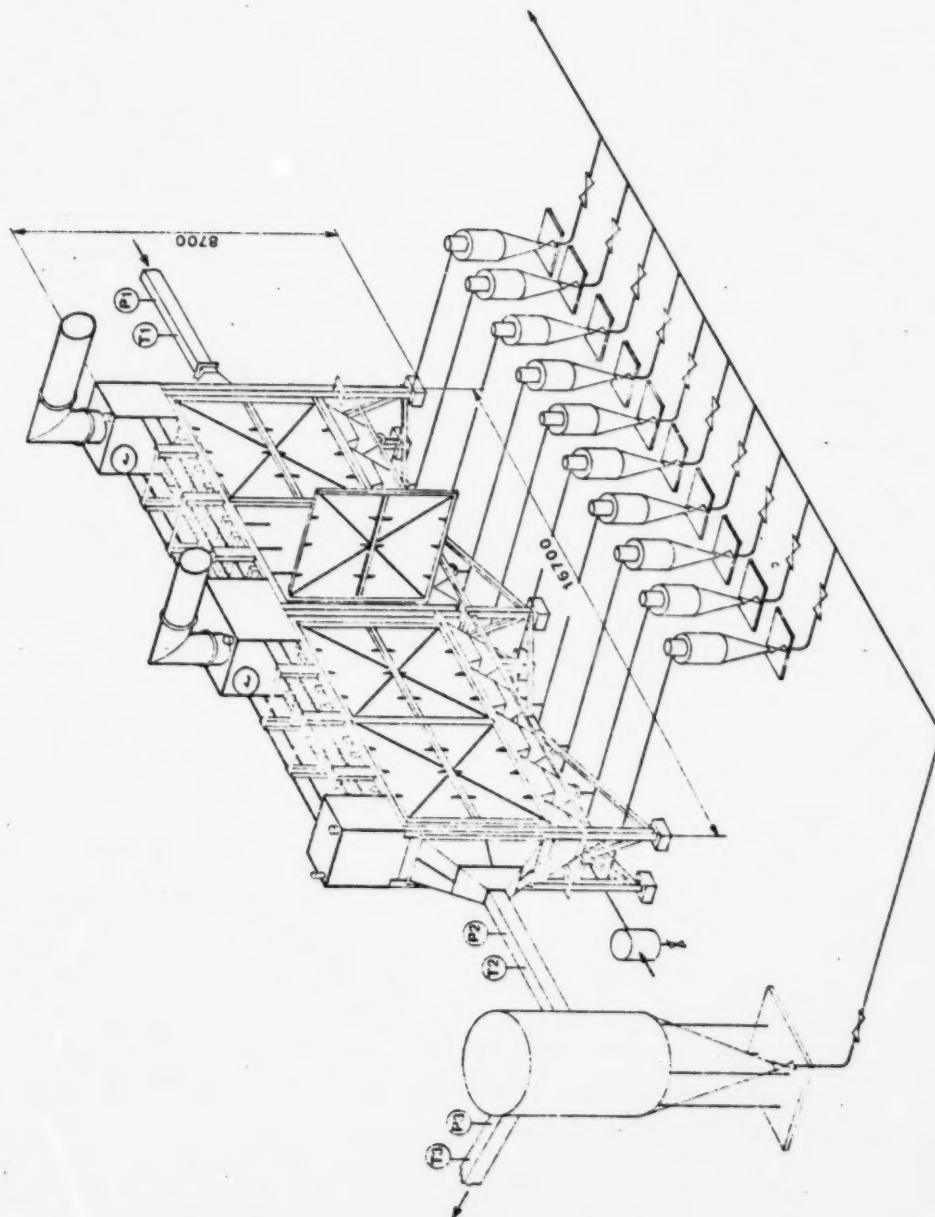


Fig. 4 : Sketch of the pilot plant (IPSE)

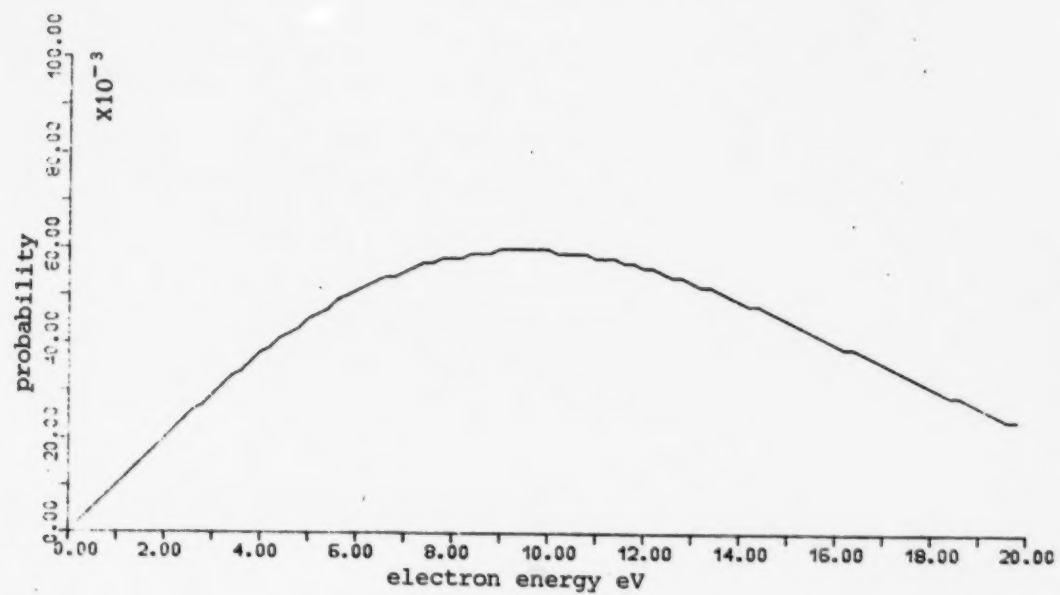


Fig. 5 : Electron energy distribution in a pulse corona discharge (Druyvenstein distribution).

UTILITY CASE HISTORY ON FIRING ENHANCED EMULSIFIED FUEL OIL
TO REDUCE PARTICULATE AND NO_x EMISSIONS

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ABSTRACT

This paper presents a recently completed test on 60-80 MW CE utility boiler firing enhanced Emulsified Fuel oil to demonstrate reduction of particulate and NO_x emissions without costly mechanical modifications. The objective of the demonstration was to ensure compliance to state and federal emission standards.

As a result of several years of basic research and development, Fuel Tech developed a patented enhanced emulsion technology which was utilized for the demonstration. This technology, coupled with specially engineered emulsifying equipment, accomplished a 26% particulate reduction and a 10 - 15% reduction in NO_x. The paper details:

1. Boiler operating data
2. Emission test data
3. Key emulsion parameters
4. Theory of enhanced emulsion firing
5. Economic considerations

INTRODUCTION

The concept of creating an emulsion with fuel oil as the external phase and water as the internal phase for the purpose of improving atomization and combustion has long been recognized as valid. Mechanical atomization produces a droplet of emulsified fuel that is broken up further upon entering the flame zone as the internal phase vaporizes and expands with sufficient violence to shatter the encapsulating oil droplet. The process is sometimes called secondary atomization.

In manufacturing emulsions, chemicals known as emulsifiers are used to control internal phase droplet size and emulsion stability. The water phase also provides a convenient means of introducing other water soluble materials into the furnace. In

this evaluation, magnesium components were added for slag modification and cold end protection. A water based combustion catalyst was also employed to promote burn-out of finer carbon particles. PP6H is Fuel Tech's designation for the chemical formulation used in this evaluation.

The testing program was designed to evaluate several of the known effects of firing emulsified fuel vs neat No. 6 oil in a utility boiler. The resulting data has been analyzed and the results are presented in this paper. The specific objective of the program was to improve particulate and NO_x emission not to evaluate boiler efficiency.

SUMMARY

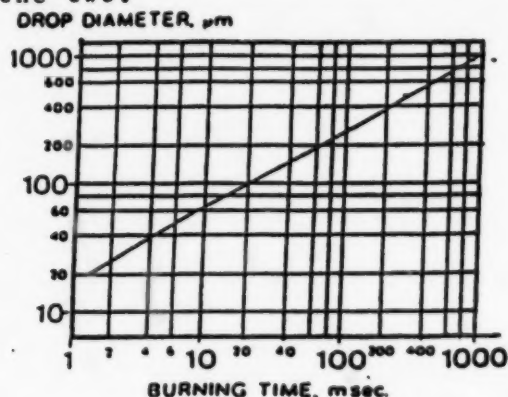
Fuel Tech Inc., in cooperation with a utility, recently completed a test program to demonstrate the effects of firing an emulsified fuel oil which was processed on site using inovative chemical and mechanical technology. Tests were conducted on a 60 - 80 gross Megawatt (GMW), CE back fired utility boiler, which had been converted from coal to oil. A medium sulfur residual fuel oil was emulsified just prior to firing with small quantities of water containing a performance enhancing chemical package.

A non load tracking emulsion system was employed and all tests were conducted at full load (60 - 80 GMW). The results of firing enhanced emulsified fuel are as follows:

- o 26% reduction in total particulate emissions
- o 10-15% reduction in thermal NO_x emissions
- o More than 20% reduction of the large particles (greater than 13.8um) in the stack
- o Increased back end acid protection. Indicated by reduced acid deposition rates measured at the air heater inlet.
- o Boiler passed state compliance tests (simultaneous particulate & NO_x) at full load operation.
- o Total water content was optimized to 7.0 - 7.5% water by volume. Mean water droplet diameter was small, in the range of 50% less than 5 um. This was caused by high shear conditions down stream of the emulsifier.
- o There was only one heat rate test performed. Although the results were unchanged within the measurable accuracy of the test, further testing should be completed before a conclusion is drawn.

THEORY

The primary reasons for incomplete carbon burnout are insufficient atomization of the oil and inadequate residence time in the flame zone. The required residence time is a direct function of the droplet size produced by the oil burner during the process of atomization. The following diagram illustrates the relationship between the two.



If the residence time of an oil drop in the combustion chamber is longer than the burning time, the drop will burn completely, but if the burning time is longer than the residence time, the drop will be extinguished when it leaves the combustion chamber and it will be emitted as a carbon particle (soot). Under the most favorable conditions, a good burner will atomize the oil to approximately 100 - 150 microns. When the conditions are not so favorable, (i.e. improper preheat, poor draft, worn tips) the burner can easily produce droplets 400 microns or larger, reducing all possibilities of complete burnout during the residence time allowed in the flame zone.

With an enhanced water in oil emulsion, combustion is improved by mechanical (micro explosion) and chemical (catalyst) means. Mechanically, by providing a secondary atomization which occurs when the dispersed water droplet, now incapsulated within the larger oil droplet, is exposed to the heat of the flame and flashes into a vapor causing a micro explosion. The larger oil droplet is shattered into smaller droplets now requiring less time to burn. Chemically, through the use of proven combustion catalysts which thermally cracks the complex long chain hydrocarbon molecules when exposed to the thermal effects in the flame. This technology is in use in refineries today. The resulting shorter chain hydrocarbons burn more quickly and completely. This combination of chemical and mechanical technologies is well suited for today's high asphaltene oils.

A more exact description of the physical processes involved when a water-oil emulsion drop undergoes combustion is as follows. It will initially be surrounded by a diffusion flame, from which heat will be transferred into the drop. The water will begin to vaporize before the bulk of the oil. The vaporization will usually occur at superheat temperatures above the normal boiling point. The degree of superheating is a function of the purity of the water, raising the boiling point 200° to 250°C with the absence of impurities.

Superheated boiling or spontaneous nucleation is known to occur "explosively"¹. A number of considerations are important to the understanding of the explosive breakup of a drop. The considerable stored energy in the superheated drop will be used partly for vaporization of the water and partly to impart kinetic energy to the oil. About 10% of stored energy is converted to kinetic energy, which is about 50 calories per gram for superheated water. If we consider a simple configuration of a 100 micron oil drop containing a 5 micron water drop under superheat conditions, then approximately 0.3 ergs will be converted to kinetic energy. This is at least 20 times as great as the surface energy of the drop, which is given by the product of the surface tension and the surface area of the drop. From energetic considerations alone, it is clear that this process has the potential to explode the drop.²

Also, by increasing the efficiency of combustion in the flame zone, less excess air is required for complete combustion. Hence, less air is available to oxidize SO_2 and nitrogen into SO_3 and NO_x .

Secondly, with the addition of the deposit and corrosion inhibitors, the catalytic action of the vanadium deposits to convert SO_2 to SO_3 is significantly reduced. In addition, what small amount of SO_3 that is formed is then neutralized.

In summary, the beneficial effects of using enhanced emulsified fuel for combustion are:

o Enhanced fuel oil droplet fragmentation through a secondary atomization ("micro-explosion" effect), and a marked reduction in ultimate fuel cenosphere size;

¹Jarvis, T.J., Donahue, M.D., and Kate, J.L., "Bubble Nucleation of Liquid Droplets Superheated in Other Liquids" Journal of Colloid and Interface Science, Vol. 50, 1975 pg 359

²Doohar, J., Genberg, R., Lippman, R., Morrone, T., Moon, S., Wright, D., "Emulsion as Fuels", ASME Winter Annual Meeting, Houston, TX, Nov. 30 - Dec. 4, 1975

- o Enhanced air utilization resulting from the increased fragmentation of the fuel oil droplets produced by the "micro-explosion" effect;
- o Enhanced carbon "burnout", with a reduction in particulate emissions.
- o The observed reduction in NO_x emissions is a direct result of the lowering combustion temperatures due to the dampening/quenching effect of the water inclusion in the fuel mixtures;
- o The water phase in the emulsified heavy fuel acts as a transport medium for introducing water soluble combustion catalysts and deposit/corrosion inhibitors.

NO_x MEASUREMENTS

NO_x measurements were taken by Fuel Tech and the Utility. Both groups employed the chemiluminescent Thermo Electron model 10 and all samples were taken on a dry basis. NO_x and NO levels were observed to be virtually the same and all values were recorded as NO_x in parts per million (ppm). Table I-III gives both raw NO_x (as recorded) and corrected NO_x (corrected to 3% O_2).

During the initial phase of the program, baseline measurements were taken with the boiler's original mechanical atomization combustion equipment in place. On August 19, because of experience in an identical boiler, coal pipes were removed and diffusers were installed making it necessary to reestablish baseline (see table II).

Fuel Tech monitored NO_x levels at the air heater inlet while the Utility monitored levels at the induced draft (ID) fan outlet ducts.

On July 16, 1985, Fuel Tech performed a 9 point baseline NO_x scan of the air heater inlet. The values were found to be relatively consistent across the duct and stratification was not considered a problem at this location.

Emulsified fuel first had the effect of dropping the NO_x values 10-25 ppm almost instantaneously. Table III shows the consistency of this effect. On August 21 & August 28, 1985, turning the emulsion process on brought the unit into NO_x compliance (less than 225 ppm corrected to 3% O_2). Measurements were made by Fuel Tech and the utility at different locations to ensure accuracy of these results. Observation of the data also indicates a reduction of excess O_2 is possible when the emulsion is fired. This effect was more pronounced at the air heater in-

TABLE I
TEST RESULTS

TEST DAY	LOAD (hp)	TYPE OF TEST	ADDITIONAL	CHEMICAL OIL RATIO	PT MEASUREMENTS				UTILITY MEASUREMENTS				COMMENTS
					TEST LOCATION	MOA (ppm) (800)	MOA (ppm) (1000)	MOA (ppm) (1500)	TEST LOCATION	MOA (ppm) (800)	MOA (ppm) (1000)	MOA (ppm) (1500)	
7-11	72	BASELINE	MoD	1:5000	NA	NA	NA	NA	10 OUTLET	215	6.3	263	ESTABLISHING BASELINE OPERATING CONDITIONS.
7-12	72	BASELINE	MoD	1:5000	NA	NA	NA	NA	10 OUTLET	225	6.5	279	ESTABLISHING BASELINE OPERATING CONDITIONS.
7-15	71	BASELINE	MoD	1:5000	NA	310	1.5	286	10 OUTLET	220	6.2	260	ESTABLISHING BASELINE OPERATING CONDITIONS, FULL BATTERY OF TESTS.
7-16	71	BASELINE	MoD	1:5000	NA	360	2.0	338	10 OUTLET	250	6.5	310	ESTABLISHING BASELINE OPERATING CONDITIONS, NOX TRAVERSE OF (PH)
7-16	71	BASELINE	MoD	1:5000	NA	340	2.4	329	10 OUTLET	NA	NA	NA	ESTABLISHING BASELINE OPERATING CONDITIONS, NOX TRAVERSE OF (PH)
7-16	71	BASELINE	MoD	1:5000	NA	320	1.2	291	10 OUTLET	NA	NA	NA	ESTABLISHING BASELINE OPERATING CONDITIONS, NOX TRAVERSE OF (PH)
7-16	71	BASELINE	MoD	1:5000	NA	330	2.3	310	NA	NA	NA	NA	ESTABLISHING BASELINE OPERATING CONDITIONS, NOX TRAVERSE OF (PH)
7-16	71	BASELINE	MoD	1:5000	NA	340	2.0	322	NA	NA	NA	NA	ESTABLISHING BASELINE OPERATING CONDITIONS, NOX TRAVERSE OF (PH)
7-16	71	BASELINE	MoD	1:5000	NA	320	3.5	329	NA	NA	NA	NA	ESTABLISHING BASELINE OPERATING CONDITIONS, NOX TRAVERSE OF (PH)
7-16	71	BASELINE	MoD	1:5000	NA	340	2.0	336	NA	NA	NA	NA	ESTABLISHING BASELINE OPERATING CONDITIONS, NOX TRAVERSE OF (PH)
7-16	71	BASELINE	MoD	1:5000	NA	340	2.9	330	NA	NA	NA	NA	ESTABLISHING BASELINE OPERATING CONDITIONS, NOX TRAVERSE OF (PH)
7-17	71	BASELINE	MoD	1:5000	NA	350	11.95	331	NA	NA	NA	NA	BASELINE CONDITIONS TESTING CONDUCTED AT THE (PH) ONLY.

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TABLE II
TEST RESULTS

FT MEASUREMENTS										UTILITY MEASUREMENTS					COMMENTS	
TEST DAY	LOAD	TYPE OF TEST	ADDITIVE	CHEMICAL:OIL RATIO	% H ₂ O	TEST LOCATION	NOx (ppm) 1 to 35	SO ₂ (ppm) 1 to 35	NOx (ppm) 1 to 35	TEST LOCATION	NOx (ppm) 1 to 35	SO ₂ (ppm) 1 to 35	NOx (ppm) 1 to 35	OPACITY R/B		
7-17	71	EMULSION	PPG4 H ₂ O	1:1000 1:5000	6.0	NA	NA	NA	NA	ID OUTLET	215	6.0	215	NA	3/8	EMULSION SYSTEM TURNED ON NOx DECREASED.
7-18	70	EMULSION	PPG4 H ₂ O	1:1000 1:5000	7.5	NA	NA	NA	NA	ID OUTLET	210	6.0	252	0.222	4/12	ESTABLISHING BASELINE OPERATING CONDITIONS.
7-22	70	BASLINE	H ₂ O	1:5000	NA	25	210	3.3	315	NA	NA	NA	NA	NA	1/2	UNIT NOT UNDER NOx CONTROL (PREDETERMINED OPERATING CONDITIONS)
7-24	70	BASLINE	H ₂ O	1:5000	NA	25	250	2.6	342	NA	NA	NA	NA	NA	8/6	UNIT UNDER NOx CONTROL
7-31	71	BASLINE	H ₂ O	1:5000	NA	25	300	3.0	300	NA	NA	NA	NA	NA	5/5	UNIT NOT UNDER NOx CONTROL
7-31	71	EMULSION	PPG4 H ₂ O	1:1000 1:5000	6.0	25	285	3.6	295	NA	NA	NA	NA	NA	5/5	UNIT NOT UNDER NOx CONTROL
8-1	71	H ₂ O OPTIMIZATION	PPG4 H ₂ O	1:1000 1:5000	6.0	25	250	3.6	259	NA	NA	NA	NA	NA	7/10	UNIT NOT UNDER NOx CONTROL.
8-1	71	H ₂ O OPTIMIZATION	PPG4 H ₂ O	1:1000 1:5000	6.0	25	240	2.2	230	NA	NA	NA	NA	NA	7/10	DETERMINE THE EFFECTS OF LOW H ₂ O. UNIT NOW UNDER NOx CONTROL.
8-2	70	BASLINE	H ₂ O	1:5000	NA	25	250	3.4	256	NA	NA	NA	NA	NA	5/6	UNIT UNDER NOx CONTROL.
8-2	70	H ₂ O OPTIMIZATION	PPG4 H ₂ O	1:1000 1:5000	9.0	25	220	2.0	210	NA	NA	NA	NA	NA	5/6	UNIT UNDER NOx CONTROL. DETERMINE THE EFFECTS OF HIGH H ₂ O.
8-20	67	REDUCED BASLINE	H ₂ O	1:5000	NA	25	210	3.7	219	ID OUTLET	195	7.2	254	NA	2/4	NEW DIFFUSERS INSTALLED TODAY PIPES REMOVED.
8-20	67	NEW DIFFUSERS EMULSION	PPG4 H ₂ O	1:1000 1:5000	9.0	25	170	4.1	181	ID OUTLET	175	7.5	223	NA	2/4	UNIT UNDER NOx CONTROL.

TABLE III
TEST RESULTS

TEST DAY	LOGD (hr)	TYPE OF TEST	ADDITIVE	CHEMICAL OIL RATIO	S-H2O	FT MEASUREMENTS				UTILITY MEASUREMENTS				COMMENTS
						TEST LOCATION	NOx (ppm) (RAW)	NOx (ppm) X02 CORRECTED (RAW)	NOx (ppm) X02 CORRECTED (170 25)	TEST LOCATION	NOx (ppm) (RAW)	NOx (ppm) X02 CORRECTED (RAW)	NOx (ppm) X02 CORRECTED (170 25)	
8-21	72	EMULSION	PPGH MgO	1:1000 1:5000	7.5	NA	NA	NA	NA	10 OUTLET	165	7.3	217	UNIT UNDER NOX CONTROL. NOX COMPLIANCE
8-21	72	BAGELINE	MgO	1:5000	NA	NA	NA	NA	NA	10 OUTLET	133	7.3	254	UNIT UNDER NOX CONTROL. UNIT OF NOX COMPLIANCE
8-22	72	EMULSION	PPGH MgO	1:1000 1:5000	6.0	25	325	3.3	315	NA	NA	NA	NA	UNIT NOT UNDER NOX CONTROL
8-28	69	COMPLIANCE	PPGH MgO	1:1000 1:5000	7.5	25	210	2.9	209	10 OUTLET	165	7.2	215	UNIT IN COMPLIANCE, FOR NOX AND PARTICULATE. NOX CONTROL PARTICULATE SIZE TEST RUN.
8-28	70	BAGELINE	MgO	1:5000	NA	25	240	3.2	243	10 OUTLET	190	7.3	232	INACH EMULSION SYSTEM TURNED OFF. FUEL FLOW READJUSTED. BOILER NOW OUT OF COMPLIANCE
9-11	72	SOOT BLOWING EVALUATION	PPGH MgO	1:1000 1:5000	7.2	NA	NA	NA	NA	10 OUTLET	165	7.2	215	SOOT BLOWING OFF. UNIT IN COMPLIANCE, FOR NOX AND PARTICULATE.
9-11	72	SOOT BLOWING EVALUATION	PPGH MgO	1:1000 1:5000	7.2	NA	NA	NA	NA	10 OUTLET	165	7.2	215	SOOT BLOWING ON. UNIT IN COMPLIANCE, FOR NOX AND PARTICULATE.
9-25	72	STATE COMPLIANCE TEST	PPGH MgO	1:1000 1:5000	6.0	NA	NA	NA	NA	10 OUTLET	165	7.0	212	SOOT BLOWING ON, MWK22 ON. BOILER IN COMPLIANCE TEST WAS OBSERVED BY THE STATE.
9-25	72	STATE COMPLIANCE TEST	PPGH MgO	1:1000 1:5000	6.0	NA	NA	NA	NA	10 OUTLET	170	7.0	219	SOOT BLOWING OFF, MWK22 ON. BOILER IN COMPLIANCE TEST WAS OBSERVED BY THE STATE.
9-25	72	STATE COMPLIANCE TEST	PPGH MgO	1:1000 1:5000	7.2	NA	NA	NA	NA	10 OUTLET	165	7.2	215	SOOT BLOWING OFF, MWK22 OFF. BOILER IN COMPLIANCE TEST WAS OBSERVED BY THE STATE.
9-26	72	STATE COMPLIANCE TEST	PPGH MgO	1:1000 1:5000	7.2	NA	NA	NA	NA	10 OUTLET	170	6.8	220	SOOT BLOWING OFF, MWK22 OFF. BOILER IN COMPLIANCE TEST WAS OBSERVED BY THE STATE.

let where there was less dilution.

In order to ensure NO_x compliance, as well as particulate, the Utility used the emulsification process to pass compliance tests at full load. In this boiler, a total H_2O content of 7% will reduce NO_x levels by 10-15%. At present this is sufficient for NO_x emissions compliance. The data indicated that higher H_2O content causes greater reductions in NO_x . However, economics relating to water vaporization costs are the limiting factors in such considerations.

PARTICULATE MEASUREMENTS

Particulate measurements were made by Fuel Tech and the utility. EPA Method 5 tests were conducted at ID fan outlet (North & South sides simultaneously) by the Utility's environmental group. Isokenetic particle sizing was conducted by Fuel Tech, on the north ID fan outlet breeching.

During the initial phase of the program, baseline measurements were taken with the boilers original mechanical atomization combustion equipment in place. On August 19, because of experience in an identical boiler, coal pipes were removed and diffusers were installed making it necessary to reestablish baseline (see table II).

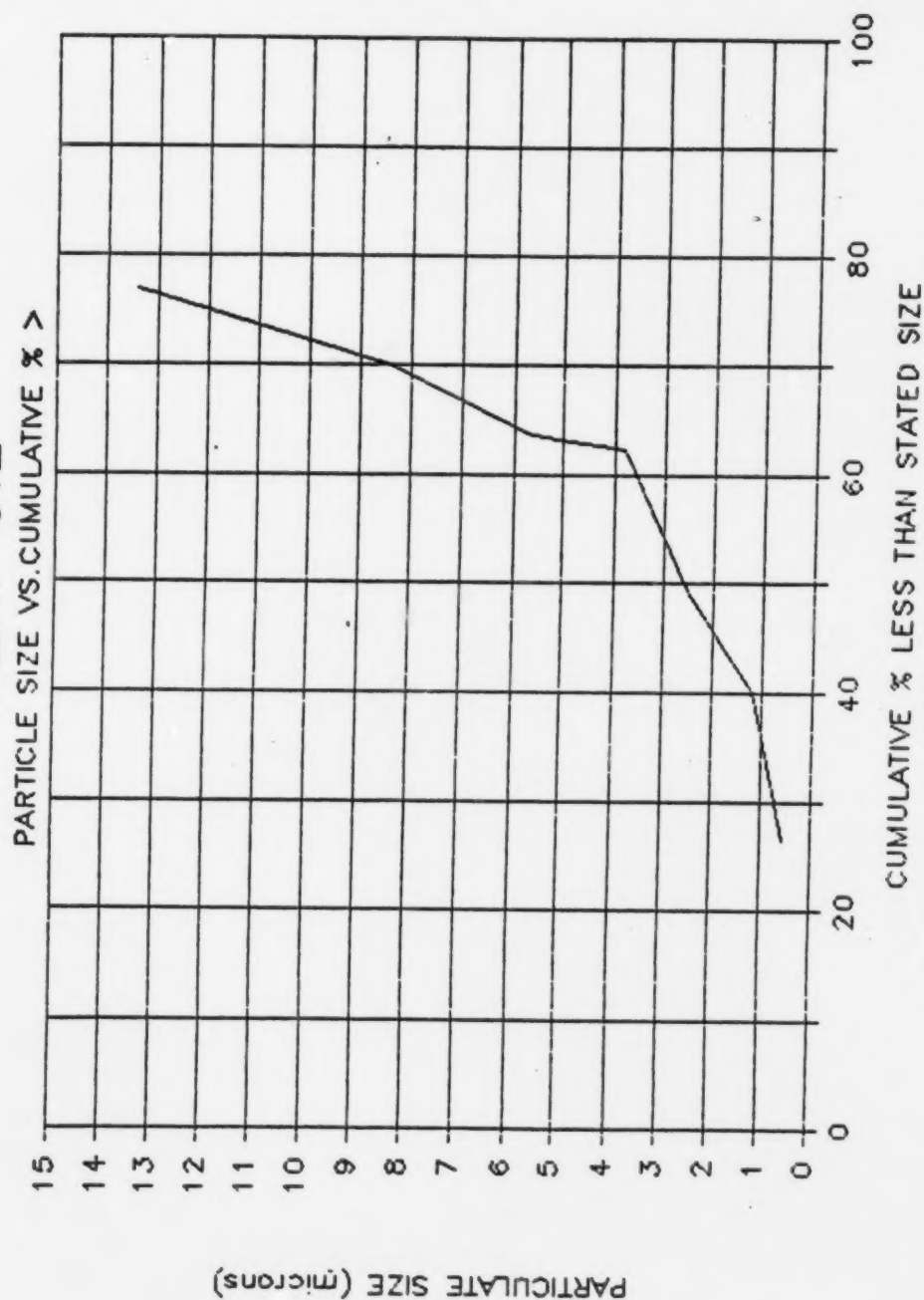
In an attempt to control as many variables as possible (fuel, operating conditions, ambient, etc.), some baselines and emulsion tests were conducted during the same day. This would allow the emulsification process system to be turned on/off while keeping operating conditions as steady as possible. Thus allowing immediate "before and after" comparisons.

EPA Method 5 tests conducted on August 28 showed a reduction in total particulate of 26% with the emulsion. The cyclones (which collect the larger particles) in the Method 5 tests also were noticeably cleaner with emulsified fuel. This indicates a reduction in amount of large particulate being emitted.

The isokenetic particle sizing tests revealed a 23% mass reduction in the large particles (greater than 13.8 μm) (see figure 1 + 2). This represents a 35% reduction in the percentage of particulates within this size range. Reduction in the amount and size of the particulate indicates a more complete carbon burn-out with the emulsified fuel.

It can be concluded from this data that the micro explosions are atomizing the oil droplets more completely, increasing the surface area and thus enabling an efficient use of the available O_2 in the flame zone. This coupled with the combustion catalyst causes a more complete combustion of the fuel.

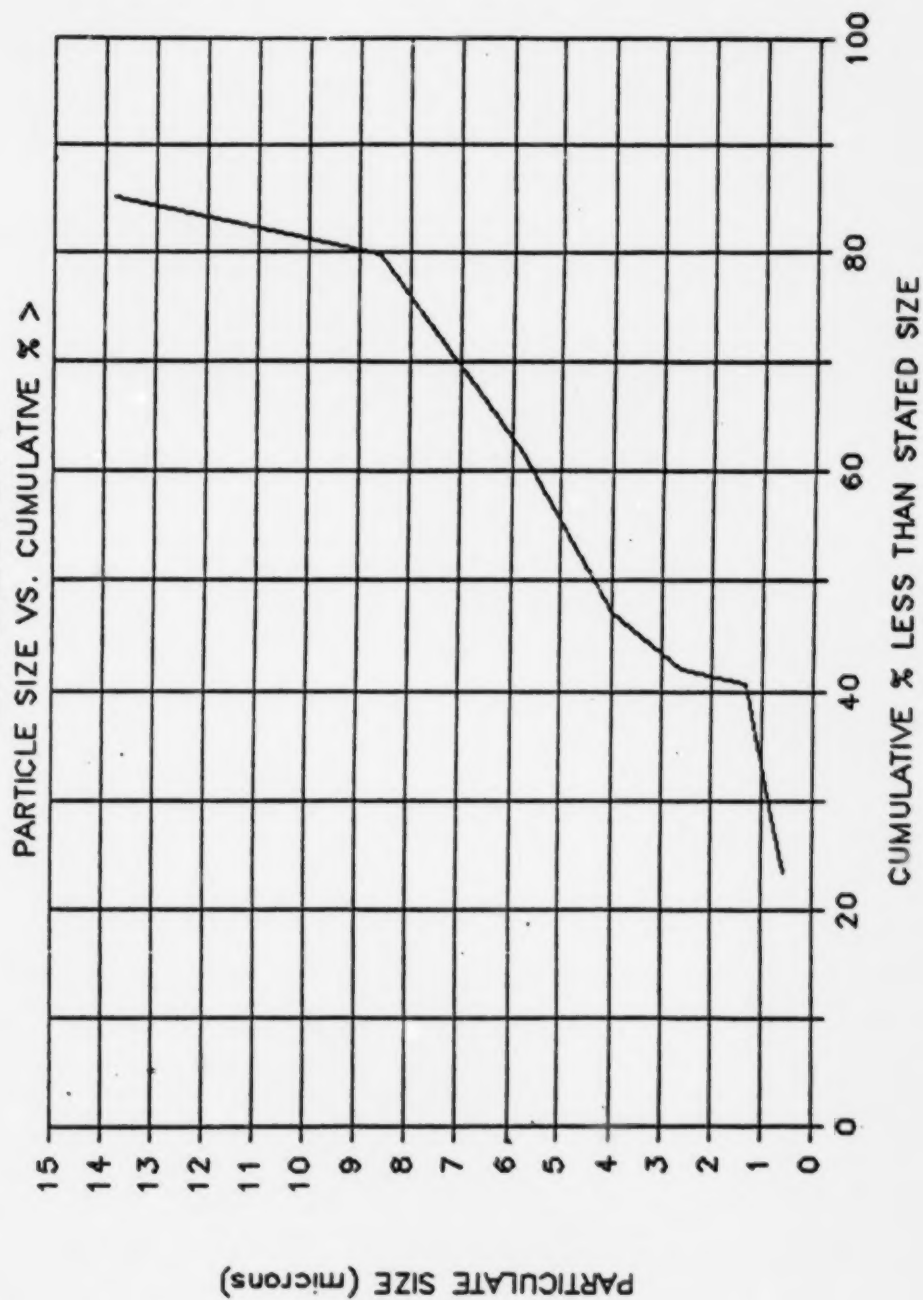
NEAT OIL



TEST 8-28-85

Fig. 1

EMULSION



TEST 8-28-85
Fig. 2

EMULSION QUALITY

The Fuel Tech emulsion consists of three individual parts, the neat oil, the water, and the catalyst enhanced emulsifying chemical (CEEC). In order to assure a high quality emulsion, each component must be precisely metered and mixed together. A specially designed burner front injection system was implemented to manufacture the emulsion on site. The emulsion was produced in the existing fuel oil line (figure 3).

Individual pumps and flow meters were used for CEEC and water. This would allow for optimization of each component. The emulsions were checked periodically to determine water content, stability and droplet size.

Mean water droplet size, at the burner, was found to be largely unaffected by emulsifier modification. This is believed to be caused by the high shear effect in the constant differential pump, a centrifical high pressure oil pump, which is located down stream of the emulsifier. As a result, mean water droplet size was less than 5 μm . Previous experience with emulsion performance indicates the optimum drop size in the range of 5 - 10 μm . Therefore, greater reductions in particulates may be possible.

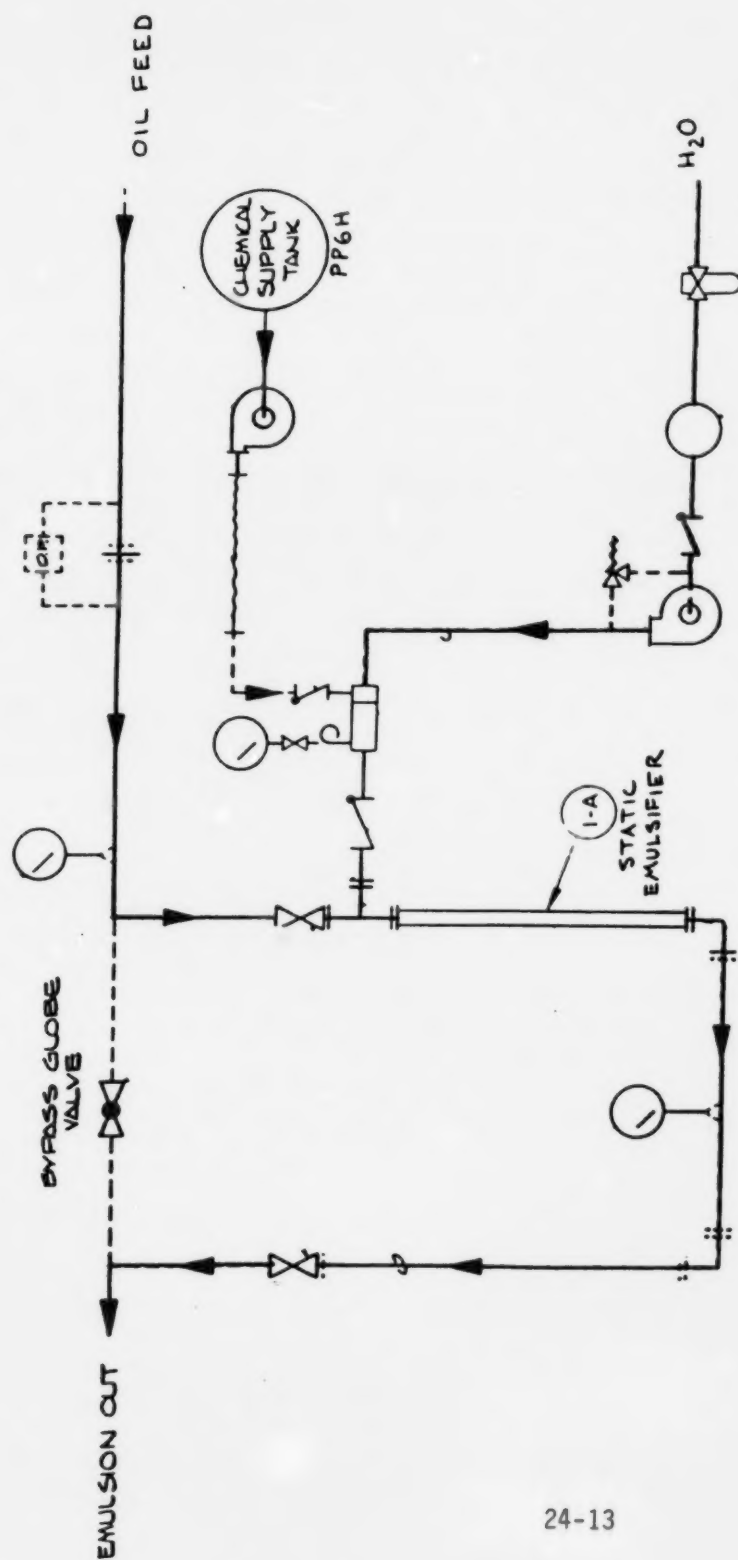
ACIDIC MEASUREMENTS

In order to monitor acidic levels in the back end of the boiler, Fuel Tech measured acid conditions at the air heater inlet. To ensure accurate results two methods were employed:

1. Severen Science Limited (SSL) SO₃ continuous monitor was used to measure gas phase SO₃ levels in-situ. The SO₃ monitor samples continuously yielding concentration in ppm (vol/vol) as a function of time.
2. Acid Deposition Rate Probe was employed to measure the rate at which acid condenses as a function of temperature, in μgm of acid deposited per cm^2 per hr. This method simulates what happens to the cold metal of the air heater baskets. Correlations between the ADR and gas phase SO₃ levels can be used to check accuracy.

GAS PHASE SO₃ RESULTS

Gas Phase SO₃ levels were monitored at six different points across the air heater inlet and the results are summarized in Table IV. Referring to the baseline SO₃ traverse on 7-16, there was SO₃ stratification. The north side was reading slightly higher than the south side. Therefore, in order to monitor the worst conditions, the majority of sampling was done



PP6H EMULSIFICATION SYSTEM
FLOW DIAGRAM
Fig. 3

on the north side.

SO₃ values ranged from 8.4 ppm (with fireside additive) to 3.7 ppm (with emulsified fuel and fireside additive). This spread is caused by different additive feed rates, firing conditions and sampling locations. This range of values indicates a satisfactory SO₃ level where back end corrosion should not present a problem.

On 7-18, baseline operating conditions were established while feeding the fireside additive and the gas phase SO₃ was measured to be 4.1 ppm @ 3.3% O₂. Then the emulsification system containing the magnesium compound was turned on and the measured SO₃ value was 4.2 ppm @ 3.1%. After approximately 1 hr, the fireside additive was turned off leaving the emulsion on. The final SO₃ value was measured to be 4.4ppm @ 3.2% O₂. This is a rise of .4ppm which is barely in the range of accuracy for the instrument.

It can be concluded from this data that the magnesium contained in the emulsion sufficiently controls gas phase SO₃ levels in the back end, thus controlling back end corrosion. It should be noted that residual effects of the previous MgO feed may bias the results and thus longer term data may vary.

ACID DEPOSITION RATE

The acid deposition rate (ADR) was measured at three different locations across the air heater inlet (see Table V) and there is a slight trend towards higher values on the north side. This would agree with the gas phase SO₃ data. Generally, there is good correlation between the ADR values and the SO₃ values.

On 7-18, (see table V) the emulsified fuel produced a greater than 50% drop in the ADR. This indicates improved neutralization of the acidic condensation on the colder metal surfaces. Therefore, we conclude that the highly reactive water based magnesium in the emulsion gives improved protection to the cold metal surfaces in the back end.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

TABLE IV
GAS PHASE SO
DATA SUMMARY

DATE	ADDITIVE	TEST PORT	LOAD (MW)	AVG % O	AVG SO (ppm)
7-10	MgO	2S	70	2.7	7.6
		2C	70	2.7	7.9
7-11	MgO	2S	72	2.9	8.4
7-12	MgO	2S	72	2.5	8.4
7-15	MgO	NORTH WALL	70	2.7	4.8
7-16	MgO	1S	71	3.5	5.5
	MgO	2S	71	2.8	5.5
	MgO	1C	71	1.2	4.9
	MgO	2C	71	2.3	5.0
	MgO	1N	71	2.0	7.0
	MgO	2N	71	2.0	5.0
7-18	MgO	NORTH WALL	71	3.3	4.1
	" + PP6H		70	3.1	4.2
	PP6H		70	3.2	4.4
7-22	MgO	NORTH WALL	70	4.1	4.3
	PP6H				
	MgO	NORTH WALL	70	4.2	5.8
7-23	MgO	NORTH WALL	70	3.1	7.1
	MgO	NORTH WALL	70	3.7	6.2
	+ PP6H				
7-24	MgO	SOUTH WALL	70	5.1	5.0
7-30	MgO	NORTH WALL	58	6.1	6.0
8-1	MgO	NORTH WALL	71	2.7	3.7
	+ PP6H				
8-2	MgO	NORTH WALL	70	4.8	4.2

NOTE: All sampling done at the Air Heater Inlet.

LEGEND: S - Southport 1 - 13" into the duct, measured
 C - Centerport from the front wall
 N - North port 2 - 39" into the duct, measured
 from the front wall

TABLE V
ACID DEPOSITION RATE (ADR)
DATA SUMMARY

DATE	ADDITIVE	TIME	TEST LOCATION	WASH ZONE TEMP. (F)	ADR (ugm/cm /hr)
7-10	MgO	9:30	AHI CENTER PORT	194 F 202 F 211 f	534 543 563
		10:30	AHI CENTER PORT	221 F 226 F 232 F	452 416 423
7-11	MgO	11:15	AHI NORTH PORT	201 F 202 F 210 F	525 507 450
7-12	MgO	9:15	AHI CENTER PORT	205 F 205 F 207 F	524 455 425
7-15	NONE	12:30	AHI SOUTH PORT	206 F 209 F 213 F	543 474 464
		13:25	AHI SOUTH PORT	222 F 225 F 232 F	455 405 425
7-18	MgO	7:58	AHI NORTH PORT	194 F 203 F 212 F	563 563 583
	PP6H	14:27	AHI NORTH PORT	203 F 205 F 209 F	198 208 217
7-24	MgO		AHI NORTH PORT	196 F 198 F 202 F	524 504 504

PARTICULATE CONTROL DEVICES FOR
MUNICIPAL REFUSE-TO-ENERGY PLANTS

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ABSTRACT

High temperature incineration followed by energy recovery in the form of process steam or electricity generation is becoming recognized as the preferred method of municipal refuse disposal for many communities in the U.S. About fifty refuse-to-energy plants are now in operation, and 20 more are under construction. Some of the existing facilities were able to comply with the Federal New Source Performance Standards (.08 grain per standard cubic foot of flue gas) without installing particulate control devices. Most of the planned facilities will use electrostatic precipitators. However, there has been a growing public concern about acid gas and organic compound emissions resulting from refuse incineration. Regulatory agencies in several states (eg. California, New Jersey, New York) now require the removal of hydrogen chloride and sulfur dioxide, and the reduction of particulate emissions to lower levels than previously allowed. Consequently, many planned incineration facilities will incorporate acid gas scrubbers upstream of the particulate control device, and some will use fabric filters instead of ESPs.

This paper will discuss the particular problems of meeting the new, more stringent emission control requirements imposed on refuse incineration. Topics to be presented include:

- . The concentrations and composition of contaminants found in incinerator flue gas, the typical fluctuations in gas conditions over time, and variations between plants.
- . Survey data on the types of particulate control

systems currently used on refuse incinerators world-wide, and actual operation and maintenance experience.

- . Specifications for a State-of-the-art combined acid gas and particulate removal system for a refuse-to-energy facility planned for New York City.
- . Impact of an upstream dry scrubber on the design and relative performance of ESP and fabric filter.

INTRODUCTION

High temperature incineration followed by energy recovery in the form of steam and/or electricity generation is becoming recognized as the preferred method of refuse disposal for many cities in the United States. Recent concerns about the environmental and public health effects associated with atmospheric emissions from municipal high temperature incinerators have prompted several states to tighten their regulation of these facilities. For example, California and New Jersey require the removal of hydrogen chloride and sulfur dioxide from incinerator flue gases, and pending legislation will impose similar requirements on planned waste-to-energy facilities in New York. There is also growing public pressure for regulations that reduce the allowable level of particulate emissions, and that control substances which have not previously been regulated, such as chlorinated hydrocarbons and heavy metals.

In recognition of these public concerns and regulatory changes, the New York Power Authority and the Energy Research and Development Authority contracted for the study described in this summary. The work was performed by the firms Malcolm Pirnie Inc. and E.T.S., Inc. during the period January 1984 through June 1985. The goal of this study was to assess the technical and economic feasibility of employing a dry scrubber followed by either a fabric filter or an electrostatic precipitator for the removal of particulate matter, hydrogen chloride (HCl) and sulfur dioxide (SO²). While other types of emission control systems are available for particulate, HCl and SO² removal, the types of systems examined in this study are the only ones being seriously considered for planned waste-to-energy facilities in the United States.

Some of the specific questions answered in this study are:

1. What are the types and concentrations of contaminants found in the flue gases of modern

waste-to-energy facilities? What range of gas conditions is expected?

2. What are the principles of operation of fabric filters, electrostatic precipitators, and dry scrubbers? What are the major components and key design features of each?
3. What modifications must be made to the design of either a fabric filter or electrostatic precipitator when the device is attached to the outlet of a dry scrubber?
4. What types of emission control systems are used in specific facilities in the U.S., Europe, and Japan? What operating and maintenance problems have been experienced, and what performance and reliability levels have been achieved?
5. What are the strictest pollutant removal levels that vendors can guarantee, and what design specifications do they recommend?
6. What are current operating and maintenance costs and total installed costs for fabric filters, electrostatic precipitators, and dry scrubbers designed to meet the stringent pollutant removal criteria found in urban areas such as New York City?

FLUE GAS CONSTITUENTS

The types and concentrations of contaminants in the flue gases of waste-to-energy facilities depend on incinerator design, refuse composition, and combustion conditions. Flue gas contaminants are typically categorized as follows:

- Particulate Matter
- Acid Gases
- Metallic Compounds
- Products of incomplete combustion

These categories overlap to a certain extent.

Particulate matter consists primarily of entrained non-combustible matter in the flue gas, as well as products of incomplete combustion that exist in solid or aerosol form. In mass burning waterwall incinerators, particulate loading at the

inlet to the emission control system ranges from 20 to 60 pounds per ton of refuse burned. Uncontrolled particulate loading expressed as concentration in the flue gas has been found to range from 0.33 to 3.5 grains per dry standard cubic foot (gr/DSCF).

Particle size distribution is an important consideration in evaluating emission control systems, especially in light of the recent emphasis on the health impacts of fine particulate in the atmosphere. Particle size distribution appears to vary widely between incinerators, and fluctuates over time at a single facility. Measurements from several plants suggest that particles smaller than 2 microns can constitute over 50 percent by weight of the total particulate matter in uncontrolled flue gas. In other measurements, however, particles smaller than 2 microns made up only 3 percent of the total.

Acid gases are the flue gas constituents which, when combined with water or water vapor, form acids. They include nitric oxide, nitrogen dioxide, sulfur dioxide and sulfur trioxide, hydrogen chloride and hydrogen fluoride. Nitric oxide and nitrogen dioxide formation in a waste-to-energy facility is a function of combustion temperature, flue gas residence time in high temperature regions, combustion air rate, and nitrogen content of the refuse. Flue gas concentrations of NO and NO² have been measured from 140 to 270 parts per million (ppm). Sulfur dioxide (SO²) and sulfur trioxide (SO₃) are formed by the oxidation of sulfur in the refuse, and therefore vary with refuse composition. Concentrations in flue gas at the inlet to pollution control systems have been found to range from under 1 ppm to 160 ppm. Hydrogen chloride (HCl) is formed by the reaction of chlorine in the refuse with hydrogen ions from the combustion of hydrocarbons. Since chlorine content of refuse varies considerably as does the fraction which reacts with hydrogen ions, the concentration of HCl in flue gas has been found to range over a broad spectrum. Uncontrolled flue gas concentrations of HCl have been found to range from 70 to over 600 ppm. Hydrogen fluoride has not been detected in waterwall incinerators in the United States.

Metallic compounds in flue gas consist primarily of oxides and chlorides of metals contained in the refuse. Most of the metallic compounds are in the vapor phase within the furnace section of an incinerator, since these compounds boil or sublime at temperatures around 1800F. The metallic compounds tend to condense as the flue gas passes through the boiler and these compounds are adsorbed onto particulate matter. A portion of the metallic compounds may still be present in the vapor phase as the gas exits the heat recovery section, but this proportion is unknown. More than a dozen metals have been detected in incinerator emissions; most are present in trace

quantities that do not cause concern. However, lead, zinc, cadmium and tin have been measured in concentrations sufficiently large to necessitate their removal by appropriate emission controls.

Products of incomplete combustion include carbon monoxide (CO), high molecular weight hydrocarbons, and chlorinated aromatic hydrocarbons. The later category, which has caused the most recent public concern, includes chlorobenzenes, chlorophenols, dibenzo-dioxins and dibenzo-furans. The mechanism of formations of the chlorinated aromatic compounds found in flue gas is still uncertain and it is being extensively investigated. About ten percent of the dibenzo-dioxin and dibenzo-furan emissions consist of compounds that are suspected of being toxic. As is the case with mettalllic compounds in flue gas, we don't know what portion of the chlorinated aromatic compounds are adsorbed into particulate matter in the flue gas. Therefore, the effectiveness of a particulate removal device in removing dibenzo-dioxins and dibenzo-furans remains uncertain.

DRY SCRUBBERS

Dry scrubbing is a process in which flue gas is made to come in contact with an alkaline material which reacts with the sulfur dioxide and hydrogen chloride in the gas and forms a dry waste product. The only commercially available type of dry scrubber used in the U.S. is a spray adsorber (also called a spray dryer), a system in which a water-based slurry of the alkali reagent is sprayed into the flue gas stream and the water is completely evaporated, cooling the gas in the process. The alkali reagent reacts with the SO_2 and HCl to form sulfate and chloride salts, most of which are carried along with other particulate matter in the flue gas to be removed by the down-stream control device (either a fabric filter or an electro-static precipitator). The most commonly used reagent is lime (calcium oxide); some spray adsorbers use nacholite or trona ores, which contain sodium carbonate and sodium bicarbonate.

The basic components of a spray adsorber are the reaction vessel and the reagent processing system. The flue gas is admitted through one end of the vessel, dispersed, and given a spiral motion by means of a set of vanes. The vessel contains an atomizer which sprays the alkali slurry into the swirling flue gas. Atomizers can be classified as either rotary or two fluid nozzle types (which use compressed air or steam to disperse the reactant slurry). The atomizer is the single largest source of forced outages in spray absorbers, largely

due to nozzle plugging or to nozzle erosion.

The reagent processing system consists of equipment that pulverizes the alkali reagent by either dry or wet grinding, and a "slaker" which mixes the reagent with water. If lime is the reagent, the slurry produced is calcium hydroxide. Since the quantity of water to be introduced per volume of flue gas is determined by the gas temperature, the concentration of reagent in the slurry is typically adjusted to provide the desired stoichiometry (i.e. molecular ratio of reagent to acid in the flue gas). The reagent feed rate is adjusted in response to acid gas concentration which can be continuously monitored either upstream or downstream of the dry scrubbing system.

In general, approximately twice as much reagent is needed to achieve 70 percent acid gas removal efficiency as for 50 percent removal. Achieving 90 percent acid gas removal would require about four times as much reagent as 50 percent removal.

FABRIC FILTERS

Fabric filters remove particulate matter from flue gas by filtering the gas through tubular bags which are sealed at one end. The bags are vertically suspended within a rigid housing and the open ends are connected to the holes in a distribution plate. As particulate matter is trapped by the fabric, a porous layer of particulate matter, called a filter cake, accumulates. This filter cake is far more efficient than the clean fabric in capturing particulate in the flue gas. However, as the filter cake thickens, resistance to gas flow increases; thus the bags must be cleaned periodically. Since bag cleaning intervals may range from several minutes to, at most, several hours, the cleaning is performed with the bags in place. Since a fabric filter system typically consists of several modular housings, each containing several dozen bags, one module is generally cleaned at a time while the others continue to filter. There are three bag cleaning methods:

1. Shaking the Bags.
2. Reversing the Gas Flow
3. Imparting a ripple motion down the bag wall by directing a pulse of compressed air at the top of the bag; this method is called pulse-jet cleaning.

In the first two methods, particulate-laden flue gas enters the inside of the bags from the bottom, and the cleaned gas is drawn from outside the bags. In the pulse-jet method, the bags are held on a rigid frame of light metal rods with closed ends

at the bottom, the particulate-laden flue gas enters from outside the bags, and the filter cake forms on the outside surface of the bags.

Two general categories of fabric used in fabric filter bags are "woven" and "felted". In woven fabric the fibers have a definite pattern; in felted fabrics the fibers are placed randomly and compressed into a dense mat. For waste-to-energy facilities, where the temperature of flue gas entering the fabric filter exceeds 300 °F, the bags may be made of fiberglass, teflon, or nomex fabric.

ELECTROSTATIC PRECIPITATORS

An electrostatic precipitator (ESP) removes particulate matter from flue gas by inducing a negative electric charge on the particulate matter and then attracting the particles to positively charged plates on which the particles are deposited. The four basic components of an ESP are: discharge electrode, collecting electrode plate, rapper and transformer rectifier.

The discharge electrodes are either weighted wires or a matrix of wires supported within a rigid frame. In the most common arrangement, the discharge electrodes are placed between the parallel vertical plates of the collection electrodes. These plates are typically spaced 20 to 30 cm apart. The discharge electrodes are energized by a high voltage power supply, typically at 30 to 50 kv DC and they ionize the flue gas as it passes around the electrode. This ionized region is called the corona discharge. Particles in this region become negatively charged. The collecting electrode plates on either side of the corona discharge are positively charged and particles are electrostatically attracted to and deposited on the plates. Collected particulate matter is periodically removed from the plates by a rapper, which imparts a mechanical impulse to the plates. The particulate falls into a hopper at the base of the ESP.

An ESP is divided into separate sections called fields, each typically 6 to 9 feet long in the direction of flue gas flow. Each field has its own voltage source and is independently controlled. Using multiple fields permits the capture of dust particles which are still entrained in the gas after it has passed through previous field(s). Adding fields increases the collection efficiency of an ESP. To satisfy the most stringent emission removal requirements, such as those imposed by New York City, outlet particulate concentration may be no greater than 0.01 gr/DSCF. Most vendors would specify a

four field ESP to satisfy this requirement. An ESP designed to meet this .01 gr/DSCF criteria would require about 40 percent more collector electrode area and would draw about 25 percent more power than an ESP designed for the same gas flow rate and a .03 gr/DSCF outlet criteria. The 0.03 criteria has been met by at least a dozen high temperature incinerators built in past five years. Most of these use three field ESPs.

RELATIVE MERITS OF ESP AND FABRIC FILTER

In designing a waste-to-energy facility, either an ESP or a fabric filter can be selected for removal of particulate matter. Each has some advantages and some drawbacks with respect to the other.

An ESP cannot be plugged by excessive particulate loading or excessive moisture in the flue gas. Fabric filter bags, however, can become irreversibly plugged due to either overloading with fine particulate matter or excessive moisture in the flue gas. This phenomenon, called "blinding", can cause flow resistance to rise above the capacity of the system's fan, thus interfering with the operation of the furnace and boiler portions of the facility.

Although metal corrosion is always a potential problem with both ESPs and fabric filters, the service life of ESP components is considerably longer than the expected life of the bags in a fabric filter. Bags typically wear out after one to three years, and bag changing is inherently labor intensive. Also, a sudden failure of many bags in a fabric filter can result from a high temperature excursion, or a spark carry over. Such incidents, however, would be far less likely to occur in a system where the fabric filter was placed down-stream of a spray adsorber. Even though the electrodes and collector plates of ESPs are subjected to gradual corrosion, ESPs are typically not prone to catastrophic failure.

The primary argument supporting the selection of an ESP instead of a fabric filter for an energy recovery incinerator is the dearth of operating experience for fabric filters in such applications, compared to ESPs. In the U.S., Europe and Japan, most large waste-to-energy facilities now in operation use ESPs. This is true both for plants with and without acid gas removal (scrubbing) systems.

Fabric filters offer several advantages over ESPs. Fabric filters, for example, have been found to be more effective than

ESPs in removing fine particulate matter, especially particles smaller than 1 micron. The removal efficiency of an ESP diminishes for sub-micron particles, while the removal efficiency of a fabric filter remains high for submicron particles.

Also, the outlet particulate concentration for a fabric filter is relatively independent of inlet particulate loading. In an ESP however, the outlet concentration is proportional to the inlet loading. Thus, if the inlet particulate loading exceeds the level for which the ESP is designed, stack emissions may not meet the desired control level.

A third advantage of fabric filters is that pollutant removal performance is not affected by the electrical resistivity of particles in the flue gas; the collection efficiency of an ESP is very sensitive to particle resistivity, which varies with temperature and particle composition.

Fourth, a fabric filter following a dry scrubber can improve the utilization of scrubber reagent. Unreacted reagent entrained in the flue gas accumulates in the filter cake and serves to continue neutralizing acid gases in the flue gas as the gas passes through the filter cake. Since flue gas does not pass through the captured fly ash on the collector plates of an ESP, this benefit is not attained in ESPs. Thus, to achieve the same degree of acid gas removal a dry scrubber followed by an ESP would require about 30 percent more reagent than a dry scrubber followed by a fabric filter.

TYPES OF EMISSION CONTROL SYSTEMS USED IN WASTE-TO-ENERGY PLANTS IN THE U.S., WESTERN EUROPE, AND JAPAN

We conducted an extensive survey of municipal waste-to-energy facilities in the U.S., Western Europe, and Japan in order to characterize the types of emission control systems in use, and to acquire data on their operation and maintenance. The survey used several methods - telephone calls to plant operators, a mail questionnaire sent to equipment vendors and facility owners, and a comprehensive literature search. We obtained information on the emission control systems of 47 facilities in the U.S., 30 in Western Europe, and 45 in Japan.

In the United States all currently operating municipal waste-to-energy facilities larger than 500 TPD use electrostatic precipitators and none of these have acid gas re-

moval systems. Two small facilities have fabric filters. These are the 180 TPD Auburn, Maine incinerator, and the 300 TPD plant in Framingham, Massachusetts. A third plant, in Gallatin, Tennessee, originally used a novel, electrostatically-charged fabric filter; however this device has since been replaced by a conventional ESP. There are also two pilot-scale fabric filters which treat a portion of the incinerator flue gas-one on the RESCO incinerator in Saugus, Massachusetts, the other on the Nashville Thermal Transfer incinerator.

At present, the only municipal waste-to-energy plant in the U.S. which has a dry scrubber is the Framingham facility. At least twelve planned facilities incorporate dry scrubbers in their designs. Six of these will be in California; the others will be in New York City (the Brooklyn Navy Yard plant), Poughkeepsie, N.Y., Bergen and Essex Counties, N.J., Waterbury and Hartford, CT. At least eight of the twelve facilities plan to use fabric filters for particulate control. The Essex and Bergen facilities plan to use ESPs.

Of the 30 European waste-to-energy facilities surveyed, nine small plants were identified in Switzerland that have fabric filters with no acid gas removal. Twenty-one facilities were identified with dry scrubbers and particulate collection systems (either ESPs or fabric filters).

Of the 47 Japanese waste-to-energy plants identified in the survey, 20 provided descriptions of emission control systems. Seventeen use dry scrubbers followed by ESPs; two have dry scrubbers with fabric filters, and one has a dry scrubber with both fabric filter and ESP.

COST COMPARISON

Budgetary price estimates were obtained for two types of emission control systems for a municipal waste-to-energy plant consisting of three 350 ton per day waterwall incinerators. One type of system was a dry scrubber followed by a fabric filter; the other was a dry scrubber followed by an ESP. Both types were required to meet the same pollutant removal criteria: 0.01 gr/DSCF of particulate, and 90 percent removal of incoming HCl and SO₂. The four vendors who supplied prices and technical specifications for both types of emission control systems priced their fabric filter systems lower than their ESPs; the differences ranged from one to nine percent. The average cost of systems with fabric filters was \$6.9 million; average cost of systems with ESPs was \$7.3 million. The dry scrubber accounted for about 60 percent of total

emission control system capital cost.

We estimated annual operation and maintenance costs for the two types of emission controls, based partly on vendor specifications and partly on data we obtained from our survey of incinerator operators. Operation and maintenance cost components included electricity, replacement parts such as filter bag, dry scrubber reagent, and manpower for operation and routine maintenance. Estimated annual O&M cost was about four percent higher for the dry scrubber/fabric filter than for the dry scrubber/ESP (\$906,000/yr vs. \$873,000/yr). For both types, the dry scrubber accounted for over 75 percent of total annual emission control system cost. To achieve 90 percent acid gas removal, about 28 pounds of lime per ton of refuse are consumed. Lime costs about 5¢ per pound.

To summarize, the cost differences between incinerator emission control systems employing fabric filters and those using ESPs are not large enough to suggest a clear preference for one type of system over the other. Rather, the choice of particulate removal system should be made on the basis of performance, reliability, and durability guarantees offered by the vendors. For example, if a fabric filter is selected, the procurement terms should include a warranty of average filter bag service life. Vendors are now willing to offer such warranties. If an ESP is selected, acceptance testing should include the sampling of outlet particulate concentration after several weeks of operation, to account for the slight degradation in performance expected during early operation.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

INTEGRATED CONTROL DEVICE FOR MICRONIC AND SUBMICRONIC
PARTICULATE AND ACIDIC GAS CONTROL

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ABSTRACT

Lone Star Steel Company has developed a family of free-jet wet scrubbers which have been proven in over a decade of operation in the steel, chemical, and wood product industries. Other successful installations include nuclear vitrification, hazardous and municipal wastes. These free-jet scrubbers remove submicronic and micronic particulate and toxic fumes to better than 99% efficiency. This has been well documented by tests sponsored and monitored by EPA and other federal as well as state regulatory agencies.

Incineration, the now accepted method for disposal of hazardous waste, is used for disposal of hazardous waste generated by government and industrial processes. Temperatures in the range of 1800 to 2000 degrees Fahrenheit are required to destruct the wastes to the 99.99 percent levels required by EPA. This creates particulates and toxic fumes (such as HCl) that are submicronic and therefore very difficult to capture in conventional gas cleaning equipment. The successful installation of Hydro-Sonic equipment on incineration processes is discussed.

Hydro-Sonic Systems is the marketing and applications division of the Lone Star Steel Company.

The work described in this paper was not funded by the U.S. Environmental Protection Agency and therefore the contents do not necessarily reflect the views of the Agency and no official endorsement should be inferred.

INTRODUCTION

Lone Star Steel has developed a family of free-jet wet scrubbers identified as "Hydro-Sonic" Cleaners. These cleaners provide performance advances in cleaning hazardous, nuclear, industrial, and municipal waste gas streams. In specific applications, economic advantages over all other methods of pollutant control are present.

The development is a by-product of wind tunnel, missile and aerospace technology. This advanced technology has been applied to gas cleaning using the technology of free-jet mixing, pressure recovery, and encapsulation employed in the development of missile propulsion systems.

THE PERFORMANCE ADVANTAGES

The performance advantages include:

1. The capture of extremely fine particulate as shown in Figure 1, from an open hearth furnace in oxygen lance where 90 percent of the particulate is less than one micron.

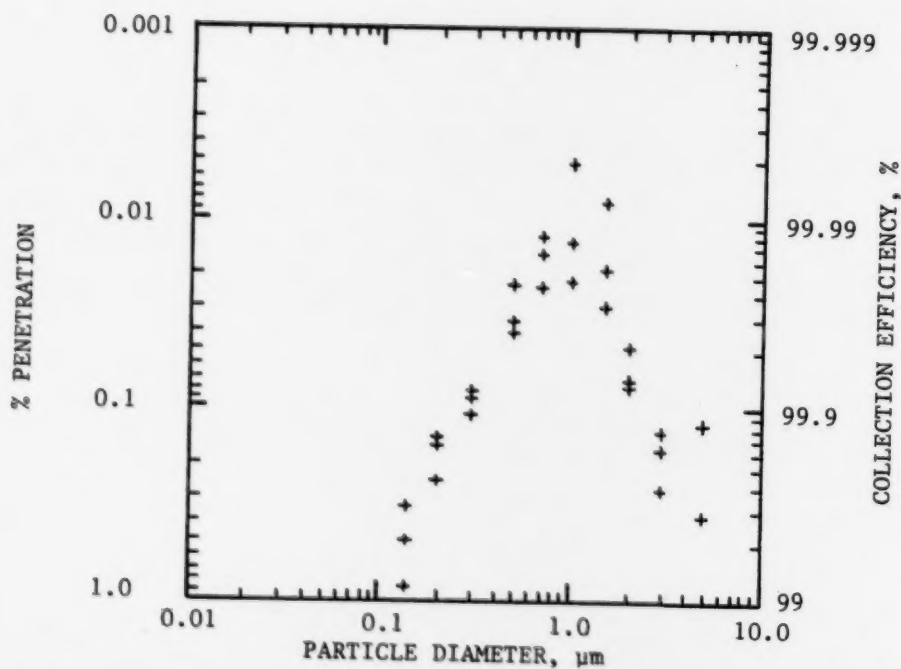


Figure 1. Fractional Efficiency Calculated Using Impactor Data

2. High level capture of acidic gases such as HCl, SO₂, and H₂S simultaneously with particulate in a single device.
3. Cleaning levels over a wide range, as required, by changing energy input but not the capital equipment.
4. Highly insensitive to inlet dust loading.
5. Instant "On-Off" in ejector models.

THE ECONOMIC ADVANTAGES

The economic advantages include:

1. Low capital cost - easy to manufacture.
2. Virtual maintenance free operation proven over a decade and a half.
3. Long life - ejector units have been in virtually continuous operation for 15 years.
4. Increased production as a by-product of "Hydro-Sonic" gas cleaning in iron and steel, pulp and paper, petrochemical, liquid and solid waste incineration, and other industrial applications by improved draft control capability and capacity.
5. Greatly reduced reagent costs over dry acid gas scrubbing processes. For example, Hydro-Sonics uses from 50 to 75 percent less reagent to capture 99 percent plus HCl as so called dry processes require to capture 90 percent.
6. Capture and recovery of fine particulate containing valuable materials in processes such as galvanizing smelters, exotic metals, and paper mills. It is well established that the fine particulate contains the greatest concentrations of minerals. The concentration can be 100 to 1000 times greater in micronic and submicronic particles over that found in the raw state.
7. Reduced use of water over conventional wet systems as a result of greater contact and mixing from free-jet actions. Free-jet expansion and mixing is the key to the easy formation of millions of droplets of the proper size and shape to contact and capture the pollutant.

BASIC CONFIGURATIONS

These cleaners are available in three basic configurations: (1) compressed fluid ejector, (2) fan driven, (3) combination of (1) and (2). These are discussed separately below. Each model provides a range of size, from under 100 cfm to above 100,000 cfm. To provide greater flexible, multiple units are often employed.

The Ejector Model shown in Figure 2 is installed in nuclear, hazardous, and industrial applications. This compressible fluid model provides extremely high performance for fine particulate and acidic gas removal and is used where it is desired to operate without moving parts in the gas stream. Pumping and cleaning are both provided without the use of fans and is attractive where abrasive, corrosive, other hostile environments are present. It is also ideal for use where waste heat is available and/or disposable wastes are burned. Highly successful applications include nuclear vitrification, hazardous, and municipal waste incineration, iron and steel making, and non-ferrous metal processes. The waste heat provides the energy to power the system in a number of applications, and in some applications provides steam for export over that required to power the cleaner.

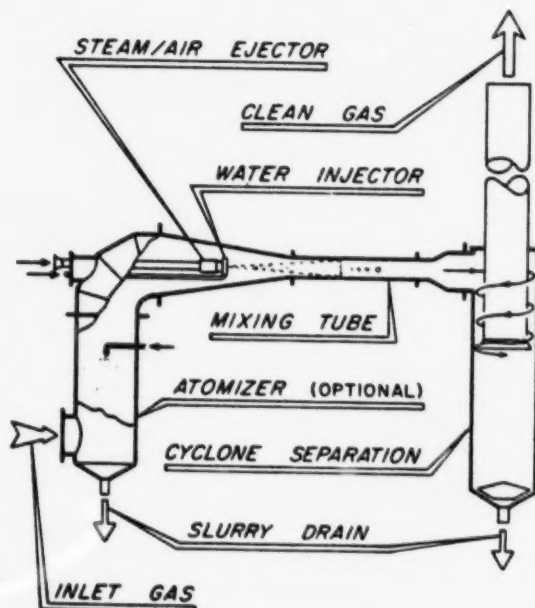


Figure 2. The Ejector Model Hydro-Sonic Cleaner.

Figure 3 shows open hearth furnaces during oxygen lance without control. Figure 4 shows the same gas stream with control using the steam ejector cleaning system. In this process 90 percent of the particles given off during oxygen lance are less than one micron. Over 40 tons of particulate are captured per day during a four furnace operation. These units are powered by steam generated from waste heat. They have been in continuous operation since 1971. Draft control provided by the system enables improved operations. These cleaners have made possible uninterrupted operation of these facilities for over 15 years.



Figure 3. Open Hearth Furnace During Oxygen Lance W/O Control.



Figure 4. Open Hearth Furnace During Oxygen Lance With Control.



The ejector model using compressed air, steam or other compressed fluid provides immediate on-off capability and is used in applications requiring this unique feature where fan and other type motive forces are not practical. An example is seen in the control of pushing emissions from coke ovens. Figure 5 is without push emission control. Figure 6 shows push emission control and the immediate "on-off" operation. As the coke push is made, the ejector models using compressed air are activated. The coke oven dust plume is drawn into the system and cleaned at the time the push is made. Cleaning continues as the coke car travels to the quenching station. A single unit provides dust control during travel. The system air supply and water supply is recharged as the coke is quenched.



Figure 5. Coke Oven Emissions Without Control.



Figure 6. Coke Oven Emissions With Control.

Principals of Operation of Ejector Models are described briefly by the following. Polluted gas is drawn into the system by the ejector nozzle which is fitted with a water injector ring as shown in Figure 7.

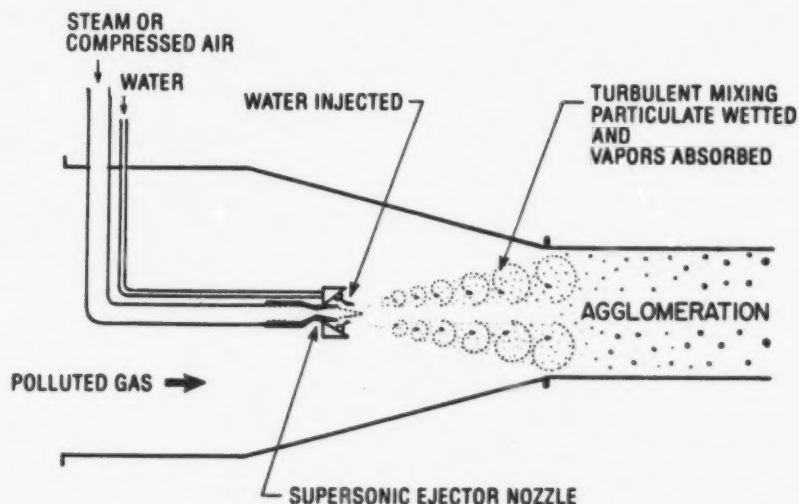


Figure 7. Ejector Drive Nozzle.

The complete wetting of the particulate is achieved initially by the generation of massive numbers of water droplets of the proper size, shape, and speed to contact the particulate and/or noxious gases. Free-jet expansion, shaping and mixing provide total contact between the pollutant and the water as illustrated in Figure 8. First, a water droplet is mechanically produced then injected onto the high speed nozzle exhaust. In the supersonic air or steam "Hydro-Sonic" cleaners the free-jet leaving the nozzle is normally in the 3000 feet per second range. When the presized water droplets strike the high speed free-jet stream, they are shattered into millions of smaller droplets of irregular shape and accelerated to high speed. It is desirable to create high-speed droplets 10 to 30 times the size of the particulate to be captured. This is made possible in the free-jet. Further, the expanding irregular shaped droplet does not permit the dust particle to follow a smooth boundary layer and therefore contact and penetration by the dust particle into the water drop is achieved with less kinetic energy employing both contact and encapsulation for capture.

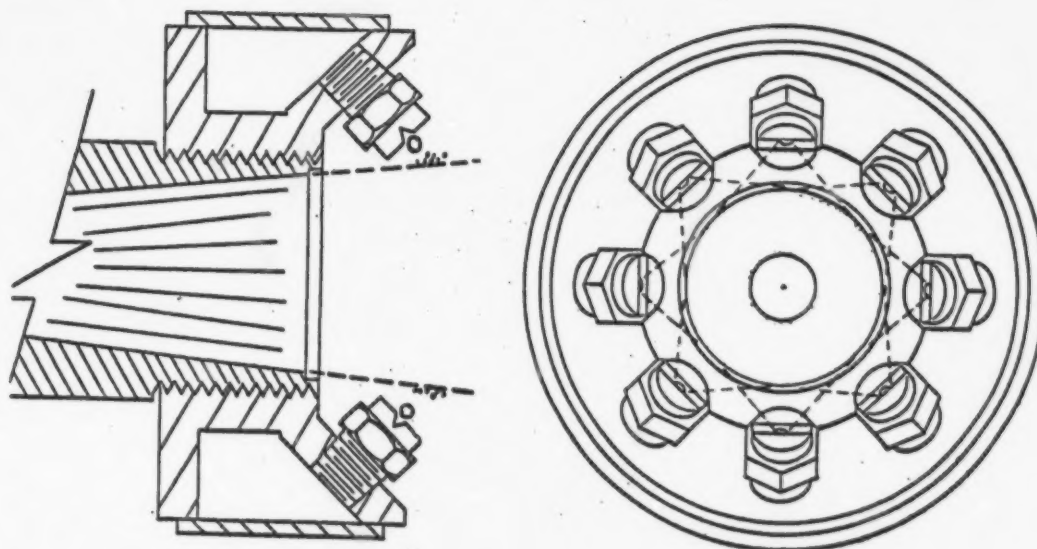


Figure 8. Water Injector Ring.

As contact is made between the water droplet and the pollutant, the droplets return to round and enslave the contacted dust particles. The mixing process occurs simultaneously with the wetting process. It insures complete wetting of any particles not already wetted and in the supersonic nozzle mixing zone is enhanced by shock waves. The shock waves form violent scrubbing zones where intimate contact is forced to occur between the scrubbing medium and the particulate and/or noxious gas and provides virtual complete capture of the pollutant.

The growth process follows and partially overlaps the mixing process so that a single water drop may contain from one to hundreds of dust particles. As a result of the growth of droplets containing particulate into increasingly larger size, the initial size and shape of the particulate in the gas stream has only a small effect on its removal.

Separation of cleaned gas and water droplets containing particulate is achieved in specially designed low pressure drop cyclones after the growth process which causes the water droplet containing the particulate to increase in size 20 to 100 times or more. These cyclones eliminate the need for other types of demisters in many cases.

The Fan Powered Free-Jet Model is employed where steam or other compressible fluid is not available and the use of a fan is possible in the gas stream. It also may be desirable to use where the gas content is high in condensibles and excess water is required.

Fan powered free-jet units (induced or forced draft) as shown in Figure 9 employ aerodynamic free-jet mixing. As with the other models, free-jet mixing sets this cleaner apart from conventional wet scrubbers. The key is free-jet expansion, droplet formation and mixing. Free jet mixing is vastly more complete and less energy intensive than confined wet scrubber mixing where numbers of drops, shapes and mixing is restricted. Therefore, higher performance is attained with less energy requirement. On some installations, where low pressure fans are installed, replacing the ineffective scrubber with a high performance Hydro-Sonic free-jet scrubber will get the facility in compliance with the more restrictive regulations without fan replacement. Hydro-Sonics uses 30 to 40 percent less ΔP than conventional wet scrubbers.

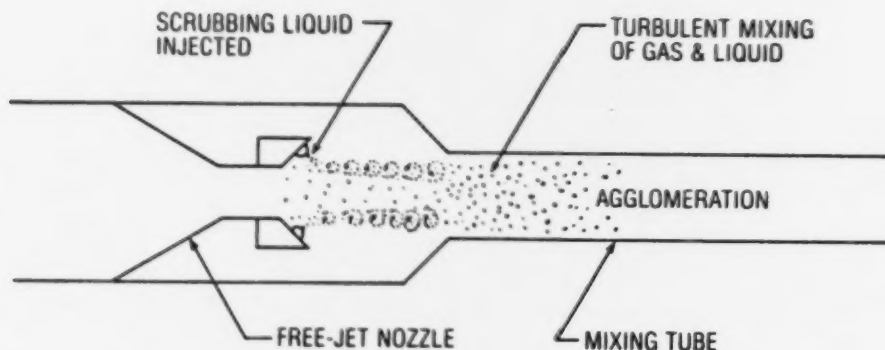


Figure 9. Free-Jet Nozzle.

Municipal waste incineration is an ideal application as shown in Figure 10. Cleaning is achieved well below regulatory requirements as shown in Figure 11. HCl removal is greater than 99.5 percent with 170 ppm incoming and less than 1 ppm out. Free-jet mixing is a superior contactor for removal of acidic gases as well as particulate on other applications. Installations consistently provide 99 percent plus removal of HCl using only 7-8 inches of water gage pressure drop.

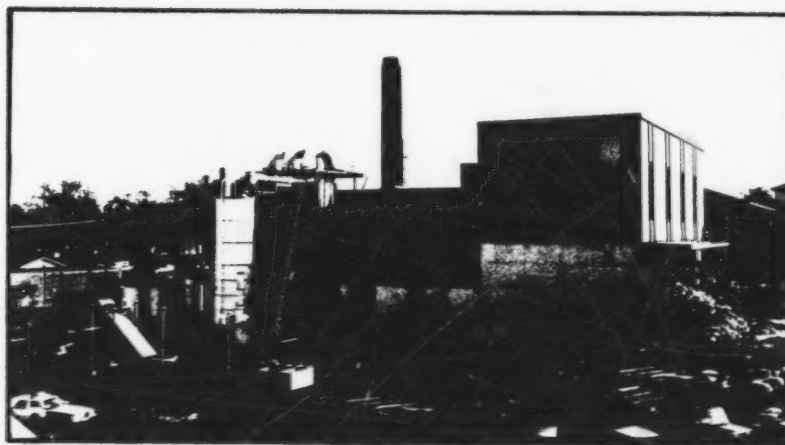


Figure 10. Shreveport Municipal Incinerator with Free-Jet Scrubber.

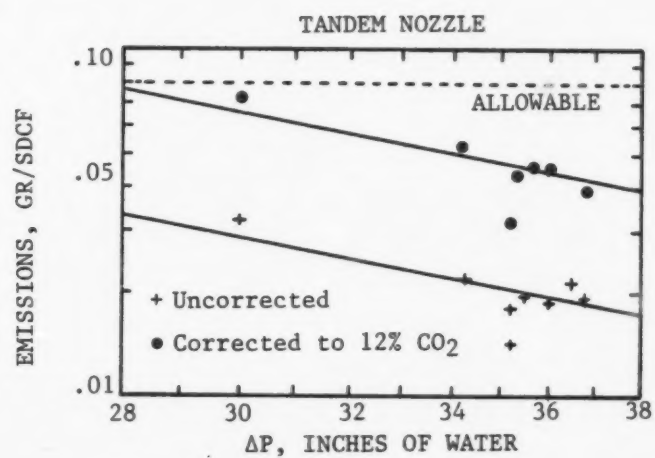


Figure 11. Municipal Incinerator Emissions Data

The Free-Jet Fan Model uses either a single nozzle or tandem nozzles determined by the inlet gas stream conditions and the efficiency desired. The tandem nozzle arrangement is recommended in condensing and removal of hydrocarbon vapors, submicronic particulate, and acidic gas. Efficiencies are significantly higher than venturi wet scrubbers. Stated conversely, the system may be operated with up to 40 percent lower energy input than conventional wet systems.

Principals of Operation of the Fan Powered Free-Jet Model.

The Tandem Nozzle Free Jet, two subsonic nozzles in series shown in Figure 12, enables this system to make maximum use of free-jet mixing and droplet sizing, vapor condensation, droplet shaping, and growth. The first free-jet section removes the large particles and initiates growth of the finest particles so that they are more easily captured in the second free-jet section. No other series (tandem) scrubber is known to show such a marked improvement in efficiency with equivalent contacting power.

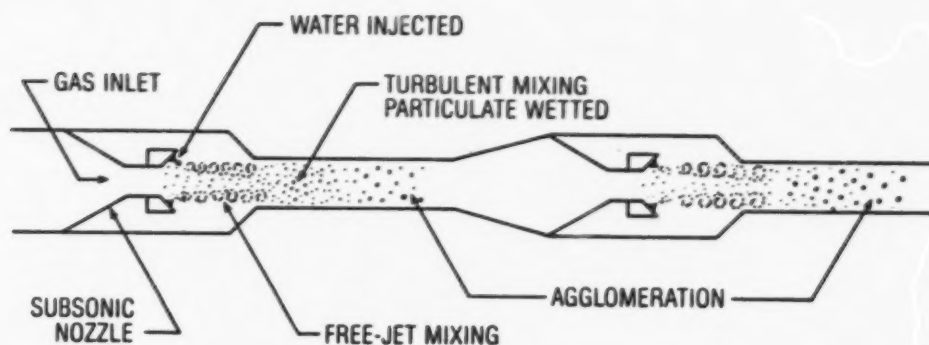


Figure 12. Tandem Nozzle Free-Jet

Particulate wetting and capture are carried out in the free-jet region. Particulate is captured when the polluted gas is mixed with fine water droplets produced by inserting water onto the exhaust of the free-jet. Acid forming gases are removed simultaneously by addition of reagents. When the water spray strikes the high speed gas (subsonic in this model) stream, it is shattered into numerous misshapen droplets in the free-jet. As contact is made with the particulate/gas, the droplet, as it slows down, returns to round (as a water drop always does). As this happens, the pollutant is encapsulated in the water drop. Impaction capture of particles is enhanced by encapsulation.

The mixing process overlaps the wetting process. It insures complete wetting of any particles not already wetted by collision or encapsulation. Mixing is accelerated and enhanced in a free-jet as compared to mixing in a confined area or region. The jet action of the emerging gas stream creates a low pressure area around the base of the free-jet nozzle causing recirculation and high turbulence which optimizes the mixing of the streams of water and gases.

The growth process follows in the mixing tube and partly overlaps the mixing process so that a single water drop may contain hundreds of micron and submicron dust particles. As a result of the growth of droplets containing particulate into increasingly larger size, the initial size and shape of the particulate in the gas stream has only a small effect on its removal. This system is extremely insensitive to high inlet grain loading. After the growth process, the large droplets containing the pollutants are removed in low pressure drop cyclones and do not require other mist eliminators. The clean gases are then vented to the atmosphere. Water and particulate are gravity drained from the cyclone bottom.

The SuperSub Fan Powered Free-Jet Model, shown in Figure 13, combines the fan driven free-jet and the ejector features previously discussed. This cleaner employs either a forced draft or induced draft fan to provide the motive force (subsonic) for the system. The use of a small supersonic free-jet ejector and water injection in front of the subsonic nozzle provides a high performance, low energy requirement arrangement to capture the finest forms of particulate.

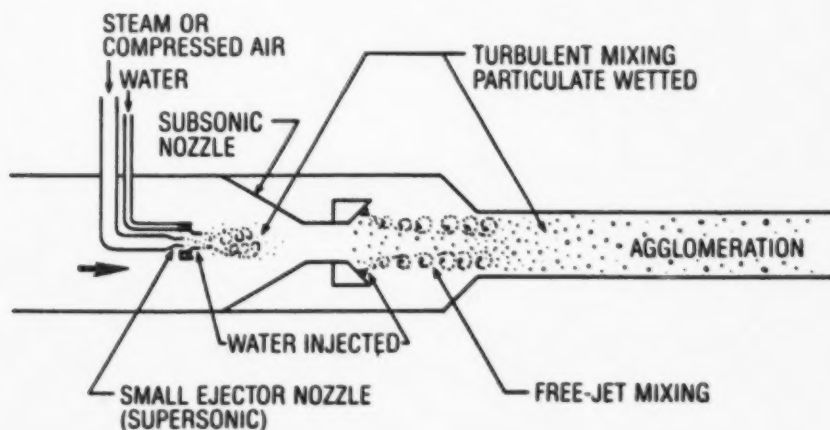


Figure 13. SuperSub Fan Powered Free-Jet Model

Use of this model is ideal where a small amount of "throw away energy" such as low pressure steam is available, and there is a need to capture submicronic particulate. Municipal waste incineration is a good application or wherever enhancement of cleaning is desired with low additional energy requirement over that attainable with a fan unit.

This supersonic addition is a low cost method of providing cleaning enhancement for extremely fine particles. It extends the range of the subsonic wet scrubbers to capture particulate into the submicron range. Air, steam, or other compressible fluids may be employed.

Principals of Operation of the Fan Powered Free-Jet SuperSub.

The SuperSub uses approximately one third of the compressible fluid in the ejector drive as does the ejector Model while lowering the fan power significantly. All of the features described in the operation of the ejector drive combine with the features of the fan drive to make this hybrid unit very attractive to upgrade a facility already having a fan in place that will not meet the requirements.

MATERIALS AND METHODS OF CONSTRUCTION

The materials of construction are selected based on the needs of the specific application. These include fiberglass reinforced plastic, corrosion resistant alloys, stainless steel and carbon steel. Operational installations range in size from 150 cfm to 365,000 cfm.

The modular design construction provides flexibility for process changes, ease of shipment, final assembly in the field, and ease of maintenance. In operation, should it be desired to take a unit "off stream", the remaining units can then be operated "off design" to keep the process at full operational level. Greater than 50 percent "turndown" is easily attained without the high maintenance cost of maintaining variable throat venturis.

CLEANING WATER REQUIREMENTS

The cleaning water requirement normally ranges from 0.3 to 2.0 pounds of water for one pound of off-gas depending on process and model selected. The cleaning water may be recirculated after clarification. In certain applications, the slurry is concentrated and used in the plant process to reclaim valuable chemicals (such as paper mill black liquor recovery boiler).

WHERE CAN HYDRO-SONIC CLEANERS BE INSTALLED

These cleaners can be installed to capture particulate, acidic gases such as HCl , SO_2 , H_2S , condensable vapors, and odors. Installations include iron & steel, petrochemical, forest products, industrial and hazardous waste, nuclear waste, and nonferrous metals.

A particularly successful application for the Steam-Hydro is on ENSCO's portable hazardous waste incinerators shown in Figure 14. Performance data is shown in Figure 15.



Figure 14. Steam-Hydro on ENSCO portable incinerator.

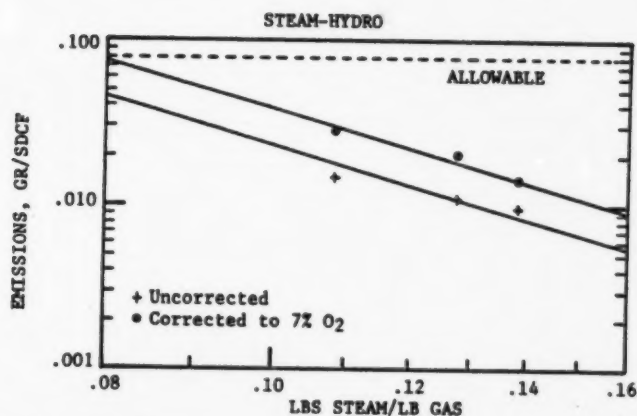


Figure 15. Portable Hazardous Waste Incinerator Performance Data

The portable hazardous waste incinerator provides cleaning well below the RCRA requirements of 0.08 gr/scf dry adjusted to 7 percent O_2 .

Figure 16 shows performance data for an industrial hazardous waste incinerator.

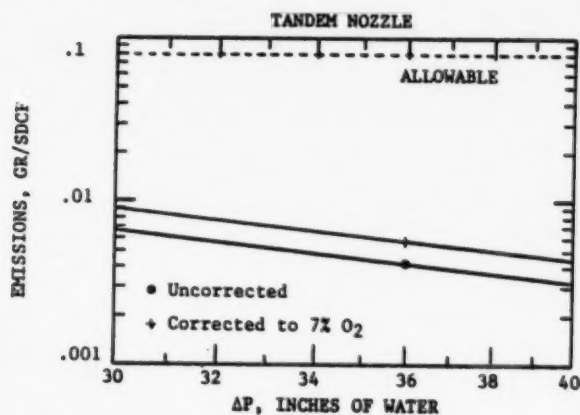


Figure 16. Industrial Hazardous Waste Incinerator Performance Data

CONCLUSION

Hydro-Sonic cleaners represent advanced technology in the air cleaning field. They provide high performance, low capital cost, and low operating costs. Maintenance is extremely low. In processes such as nuclear vitrification and burning of hazardous wastes, the Hydro-Sonic cleaners are clearly the outstanding choice. The superior performance of the Hydro-Sonic ejector model cleaner has resulted in its selection as Best Available Control Technology (BACT) at DOE Savannah River Laboratory for nuclear vitrification and control of emissions from hazardous waste. Particulate, HCl, SO₂, H₂S and other may be controlled in a single unit. The desired cleaning level is attained by changes in energy use. In the fan models, an increase in nozzle delta P provides an increase in cleaning performance. In the ejector models, an increase in the amount of steam or compressed air in relation to the off-gas being pumped and cleaned provides increased performance. The ejector driven cleaners operate from full-off to full-on in less than one second. This is a great advantage in emergency operations.

The Hydro-Sonic family of cleaners provides high performance and dependability proven in many applications. Every installation of the Hydro-Sonic cleaners has been successful covering the period 1969 to present. Hydro-Sonic Systems, Inc., (one of the Lone Star Companies) guarantees without qualification each unit sold to meet or exceed all federal and state regulatory requirements.

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